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PHYSICAL-CHEMICAL TREATMENT OF A MUNICIPAL WASTEWATER USING POWDERED CARBON

Donald E. Burns, et al

Envirotech Corporation

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Ferric chloride, alum or lime were all found to effectively produce coagulation and phosphorus insolubilization. Based on total treatment costs, including sludge disposal, alum treatment was estimated to be the economic choice for Salt Lake City municipal wastewater. Organic removal in the powdered carbon contactors was substantially enhanced by anaerobic biological activity. The use of solids-contact treatment units for carbon contacting resulted in effecting gravity clarification without the use of chemicals.

The powdered carbon physical-chemical treatment system produced a treated effluent similar to that expected for biological treatment followed by tertiary treatment for phosphorus removal. Carbon losses of 17 to 60 percent were experienced across the fluidized bed furnace regeneration system. The cause of high carbon losses was identified as ignition of carbon instead of gas which was injected into the fluidized bed to scavenge excess oxygen. (Kreissl-EPA)

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PHYSICAL-CHEMICAL TREATMENT OF A MUNICIPAL WASTEWATER USING POWDERED CARBON

By

Donald E. Burns
Gerald L. Shell

Contract No. 14-12-585
Project 17020 EFB

Project Officer

James Westrick
Environmental Protection Agency
National Environmental Research Center
Cincinnati, Ohio 45268

Prepared for

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ABSTRACT

A municipal wastewater was treated in a nominal 100 gpm pilot plant by chemical coagulation-precipitation, powdered activated carbon adsorption and granular media filtration. Spent carbon was gravity thickened, vacuum filter dewatered and thermally regenerated in a fluidized bed furnace. Solids-contact units were used for chemical treatment and carbon contacting.

Ferric chloride, alum or lime were all found to effectively produce coagulation and phosphorus insolubilization. Based on total treatment costs, including sludge disposal, alum treatment was estimated to be the economic choice for Salt Lake City municipal wastewater.

Organic removal in the powdered carbon contactors was substantially enhanced by anaerobic biological activity. The use of solids-contact treatment units for carbon contacting resulted in effecting gravity clarification without the use of chemicals.

The powdered carbon physical-chemical treatment system produced a treated effluent similar to that expected for biological treatment followed by tertiary treatment for phosphorus removal.

Carbon losses of 17 to 60 percent were experienced across the fluidized bed furnace regeneration system. The cause of high carbon losses was identified as ignition of carbon instead of gas which was injected into the fluidized bed to scavenge excess oxygen.

This report was submitted by Eimco Processing Machinery Division of Envirotech Corporation in fulfillment of Project #17020 EFB, Contract #14-12-585 under the sponsorship of the Office of Research and Monitoring, Environmental Protection Agency.

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SECTION I

CONCLUSIONS

1. The powdered activated carbon - physical-chemical treatment process evaluated produced a highly clarified effluent with COD values of 8 to 36 mg/ℓ for carbon dosages of 350 to 75 mg/ℓ.

2. Alum, ferric chloride and lime were equally effective in achieving coagulation and phosphorus insolubilization.

3. Alum treatment of 10 mgd of Salt Lake City wastewater for phosphorus removal and clarification was estimated to be less expensive than $FeCl_3$ or lime treatment.

4. The removal of soluble COD by powdered activated carbon was not significantly influenced by the type of chemical used in the chemical treatment step.

5. Gravity clarification in the carbon contactors was accomplished at overflow rates up to 0.8 gpm/sq ft without the addition of chemical flocculating agents.

6. Two-stage counter-current carbon contacting was found to require less carbon than single-stage carbon contacting to produce a given effluent soluble COD; however, the pilot plant results were not precise enough to define the difference with a significant level of statistical confidence.

7. Anaerobic biological activity in the carbon contactors resulted in removal of significant amounts of soluble organic material.

8. Laboratory equilibrium adsorption isotherm test results could not predict pilot plant soluble organic removal.

9. Alum - powdered carbon treatment of 10 mgd of Salt Lake City wastewater, with thermal regeneration and reuse of carbon, to produce an effluent containing 0.5 mg/ℓ phosphorus, 5 mg/ℓ suspended solids and 24 mg/ℓ COD was estimated to cost 18.3¢/1000 gallons. If powdered carbon were not regenerated, but used only once, the total treatment cost was estimated to be 21.5¢/1000 gallons.

SECTION II

RECOMMENDATIONS

1. Conduct a long term study using powdered carbon physical-chemical treatment process under conditions of diurnal flow to develop design, operating and efficiency data.

2. Evaluate the use of automatic process monitoring and control systems, for example:

    a. automatic organic carbon monitoring around the carbon treatment process.

    b. automatic phosphate monitoring after the chemical treatment stage.

    c. automatic density control of chemical and carbon slurry and blowdown solids.

    d. automatic coagulation control.

3. Conduct pilot plant scale gravity thickening and dewatering tests of chemical-sewage sludge to develop design data.

4. Conduct a comparative study of granular activated carbon and powdered activated carbon systems at pilot plant scale following chemical treatment of raw wastewater.

5. Conduct a thorough economic analysis of the powdered carbon process to determine the effect of plant size and effluent requirements, identifying areas of applicability. Carbon on a throw-away basis should be included.

6. Additional studies should be made to refine the powdered carbon regeneration process and obtain long-term regeneration and reuse data.

SECTION III

INTRODUCTION

Physical-Chemical Treatment (PCT) of wastewater, as discussed herein, consists of chemical coagulation-precipitation followed by activated carbon contacting. Raw wastewater is treated with chemicals and gravity settled to remove suspended solids and phosphorus. The clarified wastewater is then contacted with powdered activated carbon to remove soluble organic material. If a highly clarified effluent is desired, granular media filtration is employed.

Physical-Chemical Treatment was originally developed as a tertiary method for reclamation of secondary biological effluents. In recent years, PCT has been evaluated as a method for treatment of raw wastewaters. Because of well developed regeneration technology, major emphasis has been placed on studying granular activated carbon. The use of powdered activated carbon (PAC) has, however, several potential advantages when compared to granular carbon; namely:

1. The cost of powdered carbon per pound is 1/4 to 1/3.
2. Powdered carbon will equilibrate with soluble wastewater organics in a fraction of the time.
3. Powdered carbon is easily slurried and transported, and can be supplied on demand by metering pumps.
4. Powdered carbon might be supplied on demand to meet varying feed organic strength.
5. A powdered carbon system requires a fraction of the carbon inventory.
6. A powdered carbon adsorption system has considerably less headloss.

A recent study has indicated that thermal regeneration of powdered carbon is technically feasible and may be economical.[1] This fact coupled with the above potential advantages prompted the present study.

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The specific objectives of this study were as follows:

1. Establish operating parameters of a PAC-PCT treatment system.

2. Determine the effect of PAC dosage and methods of plant operation (i.e., single-stage versus two-stage counter-current) on removal of soluble organics.

3. Accumulate operating and process efficiency data to establish process design criteria for the production of high and moderate quality effluents.

4. Demonstrate fluidized bed thermal regeneration and reuse of powdered carbon.

5. Make an economic analysis comparing different chemical and PAC treatment approaches, establishing total and incremental treatment costs.

6. Determine if a correlation exists between accepted laboratory simulation methods and pilot plant scale data.

APPROACH

During the last six months of 1969, a nominal 100 gpm constant flow pilot plant was constructed. The powdered carbon regeneration facility was not constructed at this time.

During this same period, laboratory tests were conducted for the purpose of determining PAC and chemical requirements for producing desired effluents and for predicting plant operating conditions.

The first six month period was designated as Phase I of the study.

Phase II comprised calendar year 1970 and the first six months of 1971. Three chemicals (lime, ferric chloride, and alum) and the effect of polyelectrolytes were studied. Single-stage and two-stage counter-current adsorption systems were evaluated at six and four powdered carbon dosage levels respectively. The multiplicity of variables evaluated necessitated short plant test runs of two to six weeks.

During Phase II, design of the fluidized bed furnace for carbon regeneration was finalized and construction funds were obtained. Installation of the regeneration system was completed in March, 1971.

The last three months of the contract period involved a concentrated evaluation of the carbon regeneration subsystem.

PROCESS DESCRIPTION

A process flow diagram of the powdered carbon physical-chemical treatment process studied is shown in Figure 1. Both chemical and PAC adsorption treatment were accomplished in solids-contact treatment units similar to those which have been successfully applied for years in potable water treatment. The granular media filter was also similar to that used in water treatment applications.

The spent carbon regeneration system consists of:

1. Gravity thickening,
2. Vacuum filtration dewatering,
3. Thermal regeneration in a fluidized bed furnace,
4. Collection of regenerated PAC by venturi water scrubbers.

All these unit operations are well established in the art, but have never been used at this scale to regenerate spent carbon.

With exception that no means were provided for thickening, dewatering and ultimate disposal of the chemical-sewage sludge or effluent disinfection, the process described and tested was a complete wastewater treatment system. A photograph of the pilot plant is shown in Figure 2.

PCT PROCESS BACKGROUND

Chemical Treatment

The first major unit process in PCT is chemical treatment of raw wastewater. Normally, coarse screening, comminution, and possibly degritting would be provided prior to chemical treatment. Chemical treatment comprises the addition of chemicals to wastewater to achieve coagulation of suspended solids and insolubilization of phosphorus compounds, followed by liquid-solids separation by gravity sedimentation. Unit operations historically employed are rapid mixing for chemical dispersion and reaction (coagulation, precipitation), gentle mixing for flocculation (floc growth) and quiescent detention

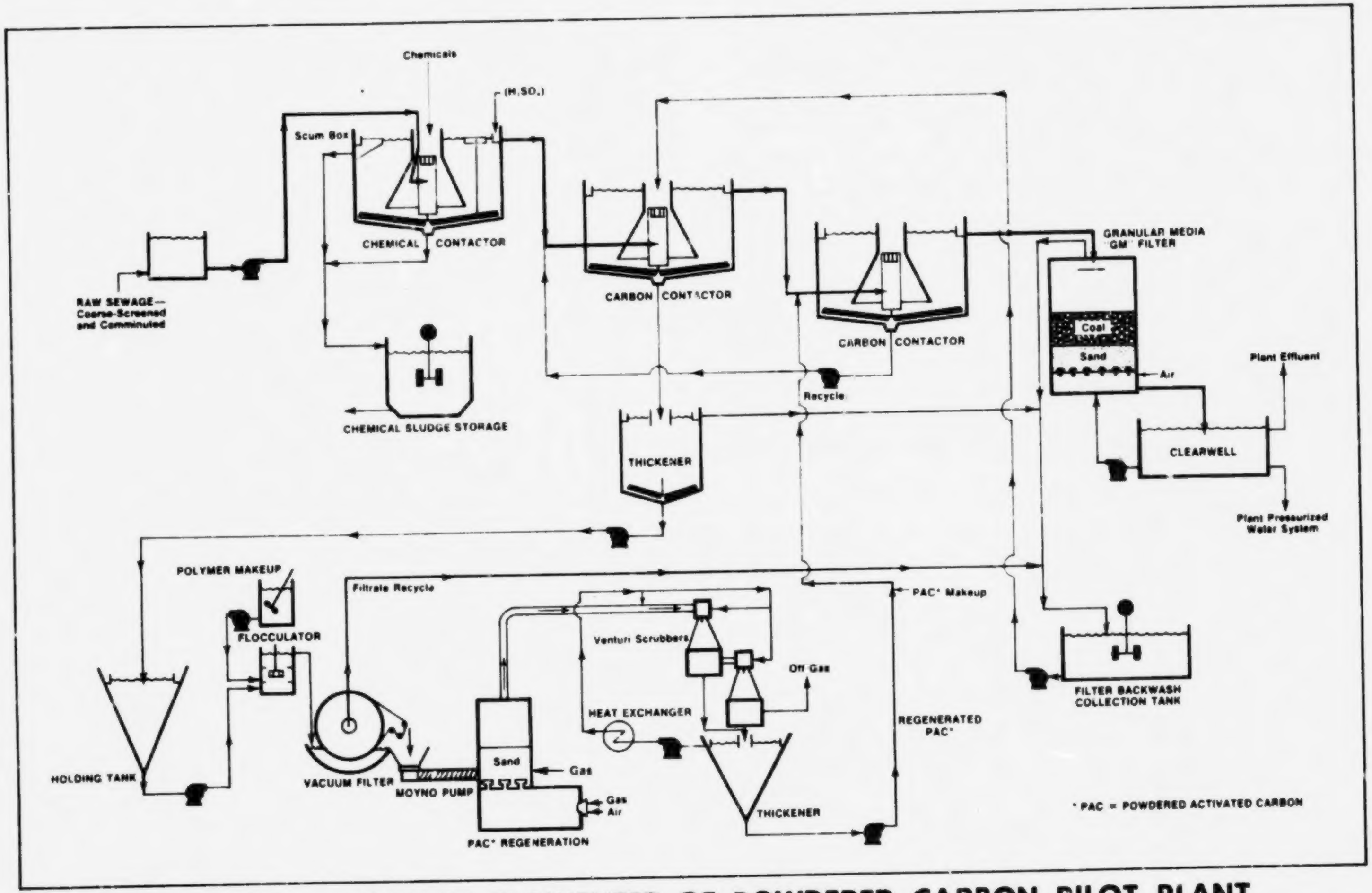
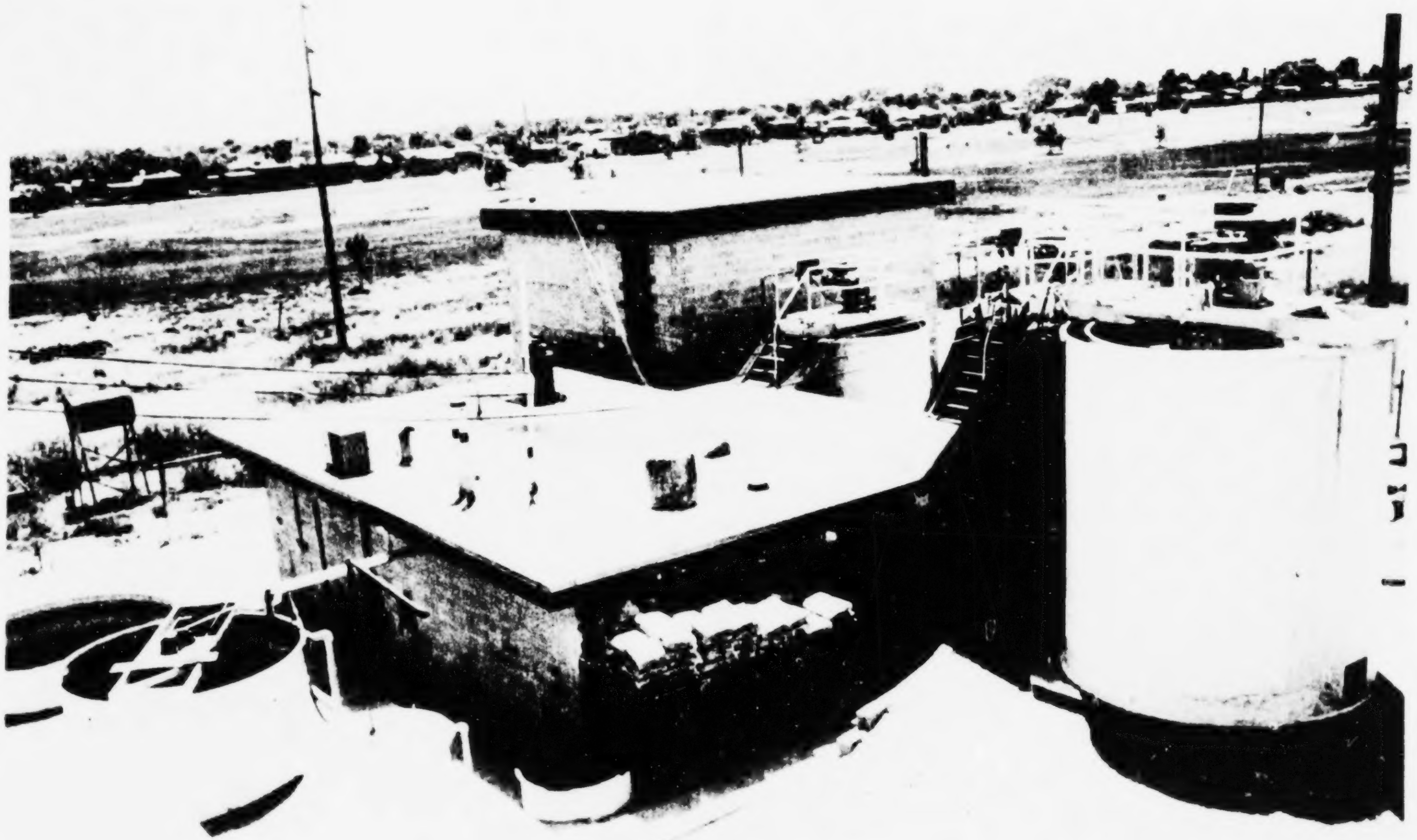

Chemicals
(H$_2$SO$_4$)
Scum Box
CHEMICAL CONTACTOR
RAW SEWAGE—Coarse-Screened and Comminuted
CHEMICAL SLUDGE STORAGE
CARBON CONTACTOR
CARBON CONTACTOR
Recycle
THICKENER
GRANULAR MEDIA "GM" FILTER
Coal
Sand
Air
Plant Effluent
CLEARWELL
Plant Pressurized Water System
POLYMER MAKEUP
FLOCCULATOR
Filtrate Recycle
Venturi Scrubbers
Off Gas
HEAT EXCHANGER
PAC* Makeup
FILTER BACKWASH COLLECTION TANK
REGENERATED PAC*
HOLDING TANK
VACUUM FILTER
MOYNO PUMP
Sand
Gas
Gas
Air
PAC* REGENERATION
THICKENER
* PAC = POWDERED ACTIVATED CARBON

FIGURE I: PROCESS FLOWSHEET OF POWDERED CARBON PILOT PLANT

FIGURE 2: OVERALL PHOTO OF PAC-PCT PILOT PLANT



for sedimentation. Considerable art on the subject of chemical treatment of wastewaters has been reviewed elsewhere[2].

A recent development for effecting chemical treatment of raw wastewater has been the use of solids-contact treatment. The interest in this approach is indicated by several pilot plant studies, for example, at Washington, D. C.; Cleveland, Ohio and Clay County, New York[3,4,5].

The key feature of solids-contact treatment units is the ability to maintain high concentrations of suspended solids in contact with the raw wastewater. Precipitated solids are accumulated within the unit to a desired quantity and kept in suspension by a submerged pumping turbine. Solids are wasted at an average rate equal to the rate of solids accumulated resulting from removal of raw wastewater suspended solids and chemical precipitates.

Coagulation of suspended solids and phosphorus precipitation by alum or ferric salt treatment are very rapid reactions, usually being completed in a matter of seconds[6,7]. Consequently, efficient rapid dispersion of alum or ferric salt throughout the raw wastewater must be accomplished prior to solids-contacting. Presumably, the benefit of solids-contacting results from improved flocculation efficiency. It has been shown that flocculation efficiency is directly proportional to mixing energy, mixing time and the number (or concentration) of particles present[8]. Solids-contact treatment provides relatively high particulate concentrations and relatively long floc detention times. For example, for chemical treatment of raw wastewater, 10-30 times as many suspended solids are maintained in contact with the wastewater in a solids-contact unit than would exist in a once through flocculation system (e.g., 0.3 - 0.9 percent suspended solids versus about 0.03 percent).

The hydraulic retention time for solids-contacting will normally be 15-30 minutes, however, the floc mean residence time will be at least several hours.

Coagulation and phosphorus insolubilization by lime treatment involves complex chemical and physical interactions. The addition of lime to wastewater simply adjusts the solution pH to a level which results in precipitation of various chemical compounds. For example, above certain pH values, phosphorus is insolubilized (e.g., hydroxylapatite), and $CaCO_3$ and $Mg(OH)_2$ precipitated. It has been reported that certain condensed phosphates are removed from solution by adsorption on chemical precipitates[9]. The coagulation-flocculation effectiveness of lime treatment presumably results from destabilization and/or agglomeration of wastewater colloids with precipitated $Mg(OH)_2$ floc. It would seem logical to expect solids-contact treatment to enhance lime treatment precipitation, adsorption and flocculation reactions.

It has been reported that lime treated wastewater is supersaturated with $CaCO_3$ and other precipitating substances; hence, there is a tendency for scale to form on equipment and pipe surfaces[10]. Application of solids-contact treatment to lime softening of water has shown that a stable water can be produced. The presence of condensed phosphates and soluble organics in raw wastewaters inhibit the growth of $CaCO_3$ precipitates. However, solids-contacting should provide more efficient reaction of lime and more complete precipitation of $CaCO_3$, compared to a once through rapid-mix, flocculation and sedimentation approach.

The presence of a large inventory of chemical-sewage solids in a solids-contact unit presents a possible problem. A sudden flow increase (greater than designed for) or chemical upset (interrupted chemical feed) could result in gross carryover of suspended solids. Proper operation will minimize this danger which also exists to a lesser degree, in a once through rapid mix flocculation and sedimentation system with or without sludge recycle.

Powdered Carbon Treatment

Appendix A presents a discussion of some academic aspects of soluble organic adsorption on activated carbon. The often used Freundlich model and the effect of counter-current contacting are discussed.

In the practical application of activated carbon to wastewater, soluble organics are removed from solution by more than just physical adsorption. The concentration of soluble biodegradable organics on the carbon surface by physical adsorption should provide a very favorable environment for biological activity. The relatively long carbon (and biomass) retention time in granular carbon columns should enhance anaerobic biological activity. In at least two studies anaerobic biological activity was considered to significantly increase organic removal compared to physical adsorption only[11,12]. Such an observation has not been reported for powdered carbon systems. The reason for this may be due to considerably shorter carbon retention times which would minimize anaerobic activity.

Conventional application of powdered carbon involves three steps: 1) rapid mixing for carbon dispersion, 2) gentle mixing for a desired contact time, and 3) liquid-solids separation by gravity sedimentation and granular media filtration.

Effective gravity sedimentation of powdered carbon suspension has required considerable chemical additions to achieve particle growth and settleability[13,14,15,16]. Alum and/or polyelectrolytes have been used for achieving coagulation and/or aiding flocculation.

Multiple stages of counter-current carbon contacting can result in significantly reduced carbon dosage requirements to obtain a given system effluent organic concentration (see example in Appendix A). One economic analysis has indicated that two-stage counter-current contacting of powdered carbon was the economic choice for treating secondary biological effluents[15]. The use of a single solids-contact treatment unit for powdered carbon treatment of secondary biological effluent has been shown to result in about one and a half stage efficiency[13]. In other words, 33 percent less carbon would have been required for solids-contact treatment compared to a conventional once through carbon contacting system. These results were later confirmed for the treatment of primary effluent[17]. In both of the above studies, it was implied that solids-contact treatment provided enhanced physical adsorption. It is, however, more reasonable to assume that biological oxidation of soluble organics caused the improved treatment results. This conjecture is based on the fact that carbon and biomass residence time in a solids-contact unit is considerably greater than in a conventional once through system.

The use of solids-contact units for powdered carbon contacting should provide an additional benefit due to "adsorptive buffering capacity". This is, the maintenance of a large inventory (mass) of carbon in contact with a time varying concentration of soluble organics results in only slight changes in system effluent. This effect can best be illustrated by an example of soluble COD (SCOD) removal on powdered carbon. Consider a constant flow, single-stage adsorption system (see Figure A-2, Appendix A), whose adsorptive relationship is described by the Freundlich model (see Curve A, Figure A-1, Appendix A):

$$\frac{C_o - C_e}{M} = (3.7 \times 10^{-5}) C_e^{2.5} \qquad (1)$$

where: C_o = feed SCOD, mg/ℓ
C_e = effluent SCOD, mg/ℓ
M = PAC dosage, mg/ℓ

The following conditions are assumed:

Flow = 100 ℓ/min
Reaction zone hydraulic retention time = 30 min
Reaction zone SS concentration = 11.5 g/ℓ of PAC
C_o = 50 mg/ℓ SCOD
C_e = 20 mg/ℓ SCOD

From Equation (1):

$$\text{Organic Loading} = 3.7 \times 10^{-5}\ (20\ \text{mg}/\ell)^{2.5} = 0.067\ \text{g SCOD/g PAC}$$

$$\text{Reaction Zone PAC} = 100\ \ell/\text{min}\ (30\ \text{min})\ 11.5\ \text{g PAC}/\ell = 34{,}500\ \text{g PAC}$$

$$\text{SCOD Adsorbed} = 34{,}500\ \text{g PAC}\ (0.067\ \text{g SCOD/g PAC}) = 2310\ \text{g SCOD}$$

If the feed SCOD (C_o) abruptly increases from 50 to 60 mg/ℓ and the PAC dosage (M) remains constant, the maximum incremental increase, after two hours (120 min) operation, in SCOD adsorbed on the PAC inventory would be as follows:

$$\begin{aligned}\Delta\ \text{SCOD adsorbed} &\simeq 120\ \text{min}\ (60\text{-}50\ \text{mg}/\ell\ \text{SCOD})(100\ \ell/\text{min})\\ &\simeq 120{,}000\ \text{mg SCOD}\\ &\simeq 120\ \text{g SCOD}\end{aligned}$$

$$\text{New Organic Loading} = \frac{2310\ \text{g SCOD} + 120\ \text{g SCOD}}{34{,}500\ \text{g PAC}} = 0.070\ \text{g SCOD/g PAC}$$

Substitution of the new organic loading into Equation (1):

New effluent SCOD = 20.5 mg/ℓ

Thus, for the example given, a 2.5 percent increase in effluent SCOD would result two hours after a 20 percent increase in feed SCOD. Had no PAC inventory been present (e.g., once through contacting versus solids-contacting), then the effluent SCOD would have increased 8.5 percent.

The solids-contact approach may provide another substantial benefit when applied to the powdered carbon process. It has been reported in the literature that suspensions of powdered carbon exhibit a self-flocculating characteristic[15,18]. In one study, concentrated suspensions were found to settle much more rapidly than dilute suspensions. This observation prompted a recommendation of sludge-blanket clarification for achieving effective liquid-solids separations of carbon slurries[15].

In another laboratory study, it was observed that following gravity sedimentation supernatant carbon concentration decreased significantly as the initial carbon concentration increased above 150 mg/ℓ[18].

As indicated earlier in this section, solids-contacting should provide improved flocculation efficiency by virtue of the high particulate concentrations and relatively long particle residence time.

Powdered Carbon Regeneration

The most promising approach to regenerating powdered carbon appears to be by thermal methods. The spent carbon is subjected to a temperature of 1200-1600 °F for a period of several seconds in an atmosphere containing less than about 1 percent oxygen by volume. Water associated with the spent carbon is vaporized and adsorbed organics are volatilized and oxidized. Proposed thermal regeneration systems require a very efficient powdered carbon off-gas separation step. Since wetting and degassing are necessary prior to reuse, wet scrubbing methods are applied to capture the regenerated carbon.

There have been at least four approaches evaluated for thermal regeneration of powdered carbon:

1. Indirect heated transport reactors

2. Partial wet air oxidation

3. Direct heated transport reactor

4. Fluidized bed furnace

No public disclosures of indirect heated transport reactors are available. In private communications, representatives of one firm indicate carbon losses across their entire full scale regeneration system of less than 5 percent[19]. The regenerated product is used in a chemical processing system.

The wet air oxidation system involves subjecting a thickened powdered carbon slurry (about 6-8 percent solids by weight_ to detention in a pressurized, steam heated reactor[20]. The reactor product (carbon slurry) is passed through a decant tank prior to reuse. Results of a laboratory evaluation indicated effective recovery of carbon adsorption properties.

The direct heated transport reactor approach involves feeding thickened spent PAC (about 10 percent solids by weight) into a direct flame reactor in which the temperature is maintained at about 1400-1600°F[21]. The carbon detention time in the reactor is approximately one second. The carbon is captured in water and air cooled jet condensers. Evaluation of a 4 lb/hr (dry carbon feed unit indicated recoveries of 90 percent and near complete recovery of adsorption characteristics[21]. Another evaluation of the same direct heated transport reactor indicated average PAC recoveries of 84 percent[1]. In this latter study, considerable difficulty was encountered in maintaining a "choke" ring in the gas burning chamber.

The fluidized bed furnace approach was developed by Battelle Memorial Institute[22]. It involves injection of dewatered, spent powdered carbon (about 25 percent solids by weight) directly into a fluidized bed of inert material (sand). The fluidized bed is maintained at 1200-1500°F by combustion gas flow. Excess oxygen in the exhaust gases is held at less than 1 percent by volume. The carbon detention time is about 4-8 seconds. Evaluation of a 4 lb/hr (dry carbon feed) unit indicated recoveries of 86 percent[1]. Operation of the fluidized bed furnace system was considered successful with no major difficulties. A comparative evaluation of the above direct heated transport reactor and fluidized bed furnace resulted in the latter being chosen for development at a larger scale[1].

SECTION IV

LABORATORY STUDY

Laboratory studies were conducted during Phase I to develop background data on the response of Salt Lake City raw wastewater to the physical-chemical unit processes embodied in the pilot plant.

Analytical and testing procedures used are presented in Appendix B. Of basic importance is the definition of soluble pollutants as those that pass through a 0.45 micron membrane filter.

CHEMICAL TREATMENT

Three chemical treatments which are generally considered to provide both coagulation and phosphorus precipitation, were evaluated by jar tests of several raw wastewater samples. The chemicals used were alum, $FeCl_3$ and hydrated lime. Figures 3, 4 and 5 show typical data from these tests. Note that soluble COD (SCOD) and phosphorus, are shown.

Alum treatment data in Figure 3 indicates a drop in pH for increasing alum dosage, as expected. An optimum alum dosage based on phosphorus precipitation and turbidity reduction, occurs at 20 mg/ℓ Al^{+3}. Above this dosage, residual turbidity increases, indicating a redispersion of colloidal aluminum precipitate (presumably a hydroxide). Soluble phosphorus remains very low, indicating that the phosphorus precipitate is not solubilized.

Of interest is the removal of some 10-20 percent of the SCOD. The significance of this observation is that a reduced amount of SCOD would report to the subsequent carbon adsorption step. Consequently, a proportionate reduction in carbon requirements would be expected.

The ferric chloride treatment results, shown in Figure 4 follow the same general trends as for alum, except that redispersion was not observed for $FeCl_3$ dosages up to 135 mg/ℓ of Fe^{+3}. The optimum Fe^{+3} dosage appears to be about 65 mg/ℓ.

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FIGURE 3: ALUM TREATMENT OF RAW WASTEWATER

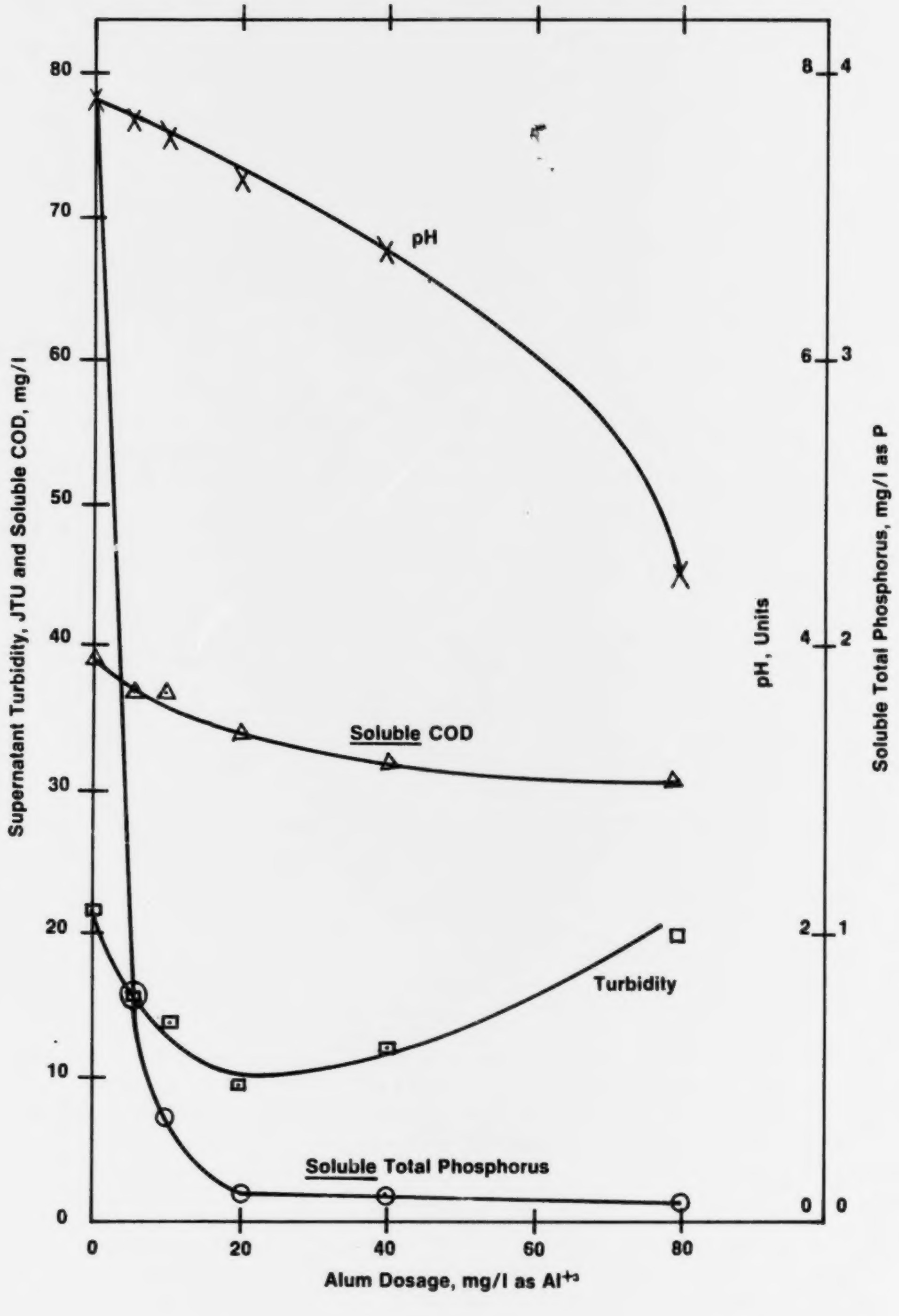


FIGURE 4: $FeCl_3$ TREATMENT OF RAW WASTEWATER

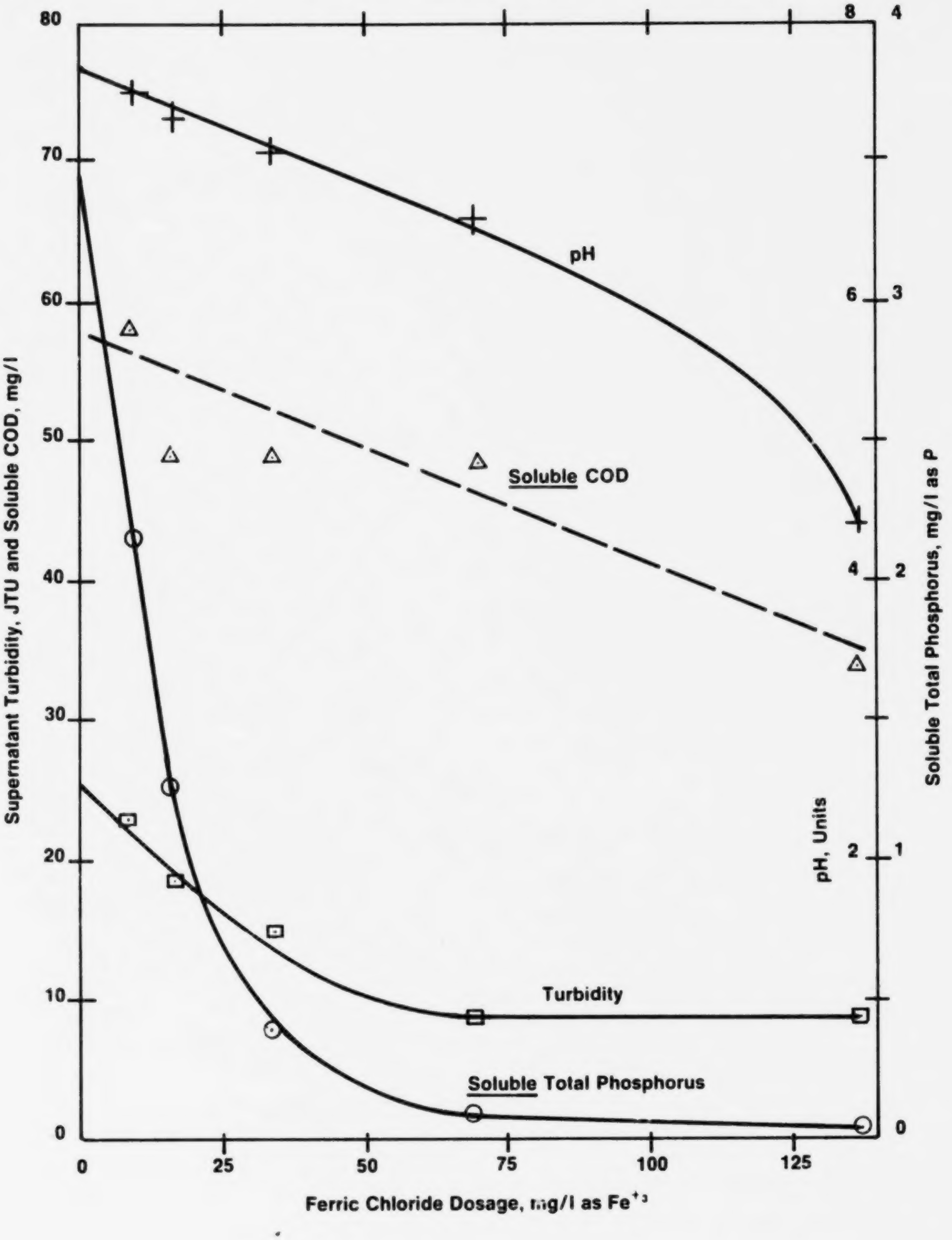
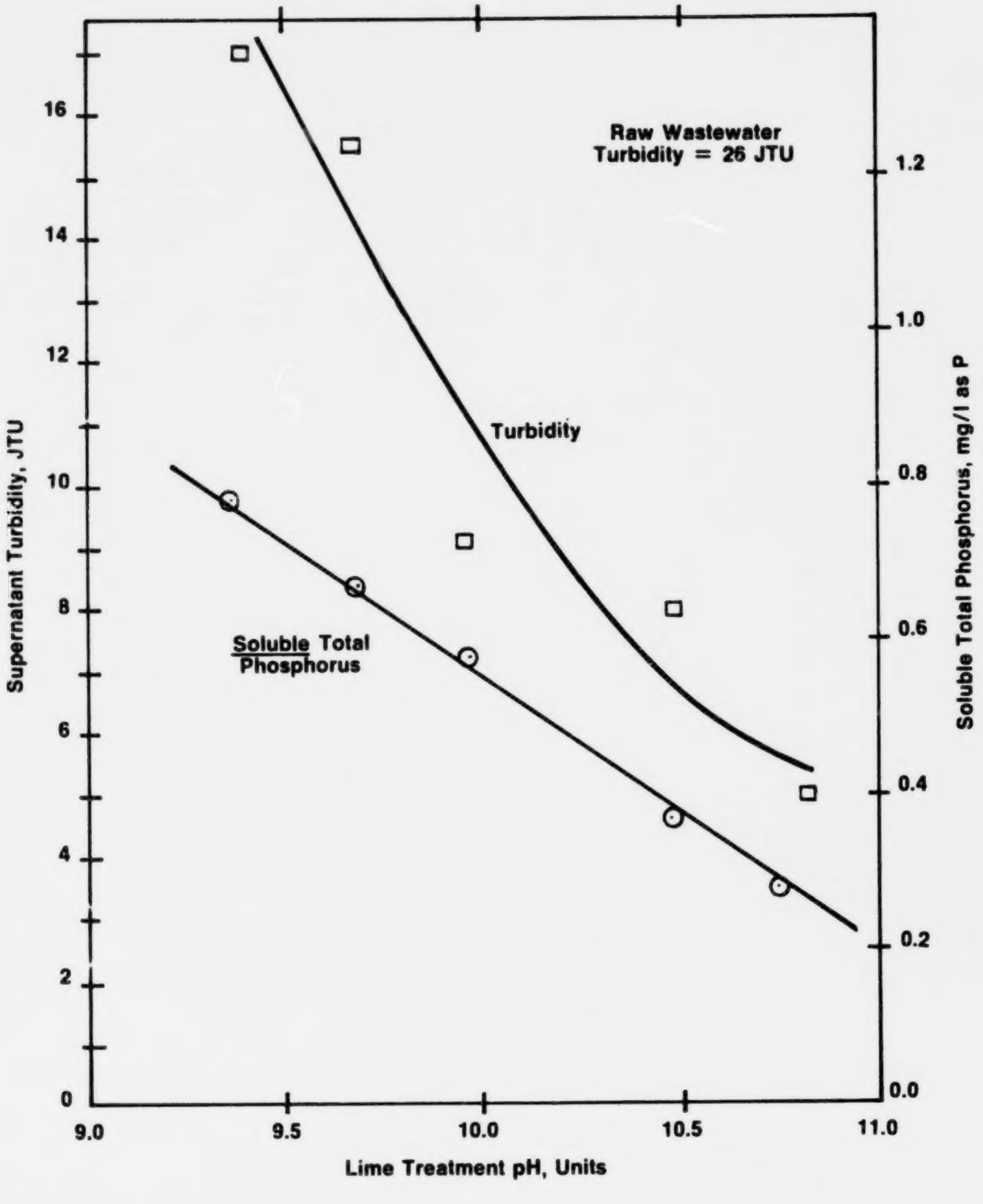


FIGURE 5: LIME TREATMENT OF RAW WASTEWATER



It appears that Al^{+3} is more effective than Fe^{+3} based on the optimum dosages noted above (e.g., 0.74 millimoles Al^{+3} versus 1.2 millimoles Fe^{+3}). Assuming filter alum)9.1 percent Al^{+3} by weight) costs 3.5¢/lb and $FeCl_3$ costs 5¢/lb, the above optimum dosages result in $FeCl_3$ chemical costs being 22 percent more than for alum.

The response of raw wastewater to hydrated lime, $Ca(OH)_2$, treatment is indicated in Figures 5 through 8. As lime is added to and reacts with the wastewater, the solution pH increases as indicated in Figure 6. One reaction occurring is the conversion of biocarbonate to carbonate alkalinity. This reaction is essentially complete at a pH of about 9. The amount of bicarbonate alkalinity initially present affected the lime dosage required to obtain a given treatment pH (see Figure 6).

Figure 7 indicates that as the pH is raised above 9.5, calcium is precipitated (as $CaCO_3$) and non-calcium hardness cations begin to precipitate (e.g., $Mg(OH)_2$). For the wastewater samples tested, minimum $CaCO_3$ solubility appears to be at a pH of 10.4. Magnesium hydroxide is generally considered to be the chemical species effecting coagulation for lime treatment. An operating pH in excess of about 10.5 is required to provide precipitation of significant quantities of magnesium hydroxide. From data presented in Figure 5, a turbidity reduction of 90 percent is not achieved until a treatment pH in excess of about 10.7 is reached. The soluble total phosphorus is reduced to about 0.3 mg/ℓ at pH 10.7.

An example of solids production resulting from lime treatment is shown in Figure 8. Solids production in the pH range of 9.5 to 10.5 is mainly due to precipitation of $CaCO_3$. In the pH range 10.5 to 11.5 few additional solids are produced, presumably only small quantities of $Mg(OH)_2$. Solids production in this pH range is about 7000 lb/MG (840 mg/ℓ). Above a pH of 11.5 solids production can only be explained as being due to unreacted $Ca(OH)_2$. As the pH increases from 10.4 to 12.5, the ratio of pounds of solids produced per pound of lime added is seen to decrease from 1.7 to 0.7.

Based on typical jar test results presented, the following approximate dosages of chemicals tested would be required to provide effective clarification (<10 JTU turbidity) and soluble total phosphorus removal (<0.3 mg/ℓ):

FIGURE 6: LIME TREATMENT: EFFECT OF INITIAL ALKALINITY

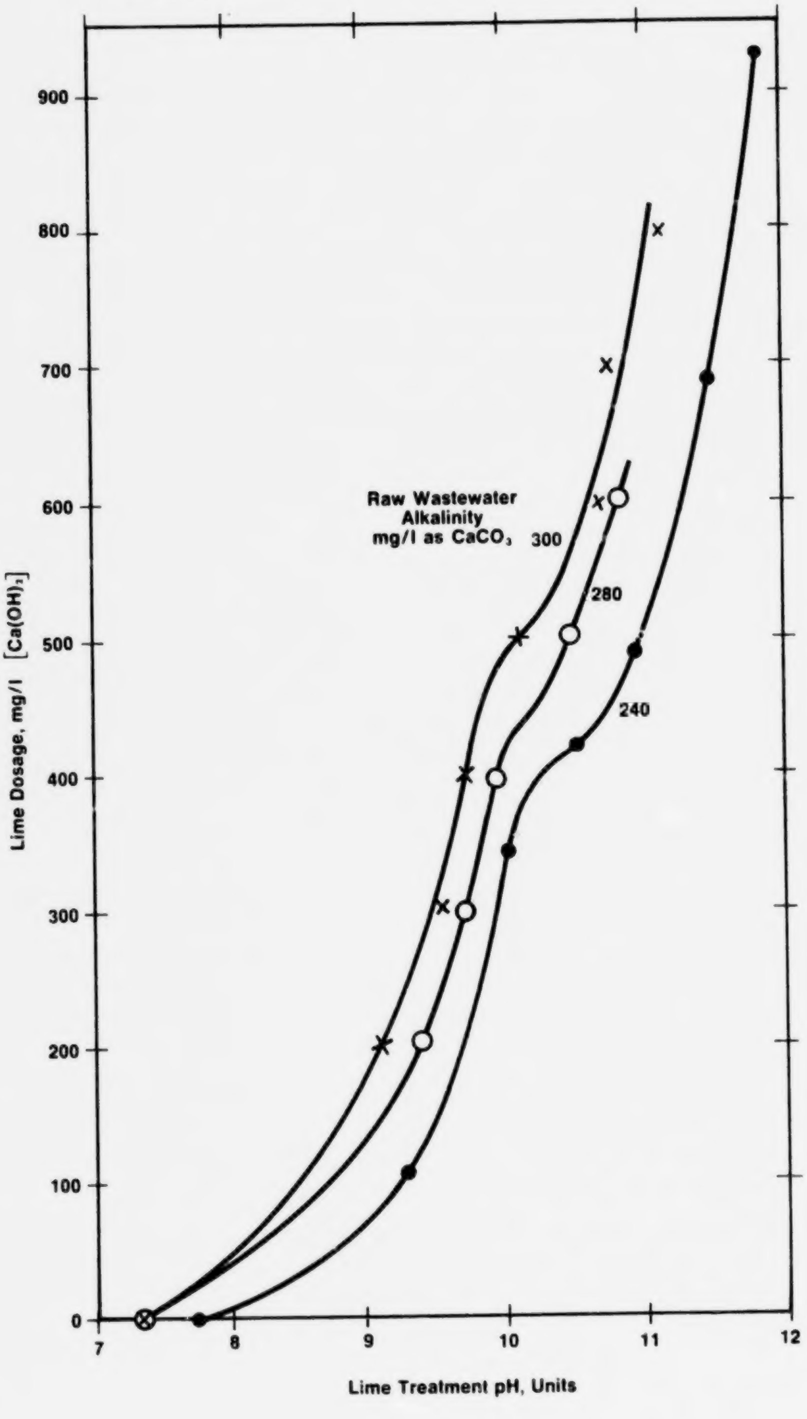


FIGURE 7: LIME TREATMENT: EFFECT ON HARDNESS

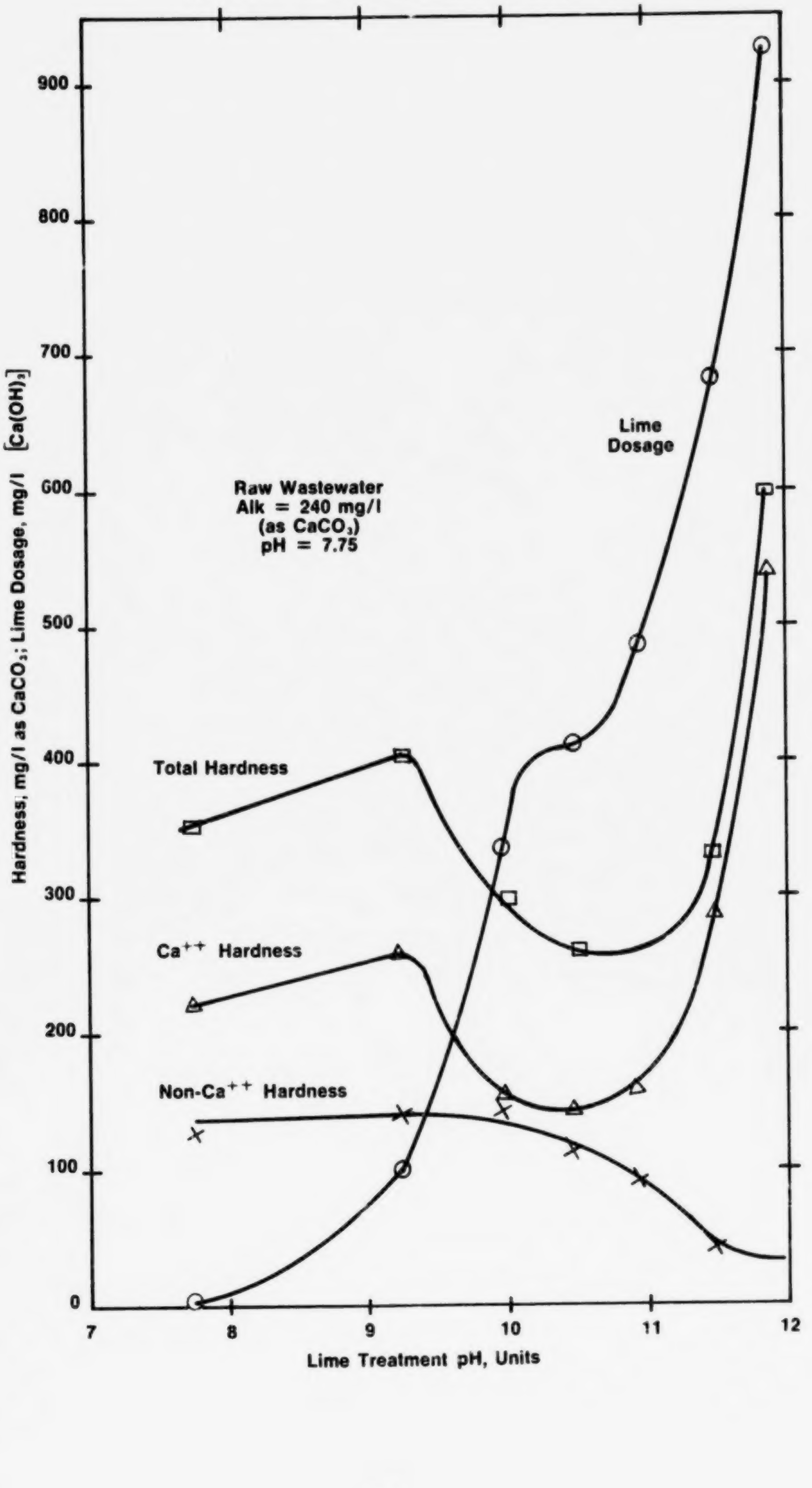
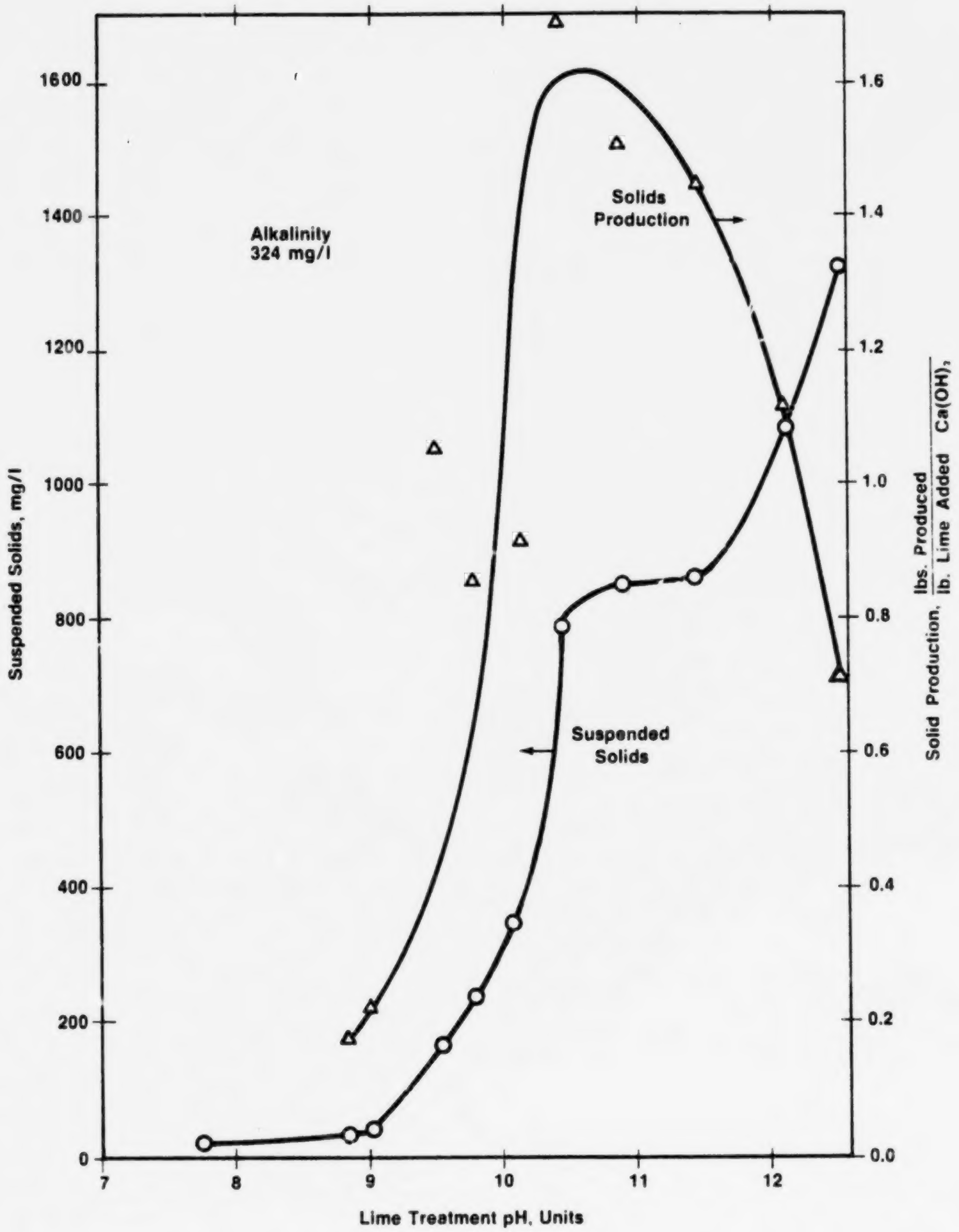


FIGURE 8: LIME TREATMENT : SOLIDS PRODUCTION



Chemical	Dosage
Alum	150 mg/ℓ
$FeCl_3$	150 mg/ℓ
$Ca(OH)_2$	450-600 mg/ℓ

These approximate dosages were used in subsequent pilot plant tests.

Since a solids-contact unit was to be used in pilot plant tests, an attempt was made to simulate this process by modified jar tests (see Appendix B for the procedure). Tests were conducted with 200 mg/ℓ of $FeCl_3$, 220 mg/ℓ filter alum and lime treatment at a pH of 11. Several tests were conducted with 4 to 6 contacts and one lime treatment test was carried through 20 contacts. The contacts indicate the number of times the settled sludge is accumulated from a single jar test. No significant change in supernatant turbidity, total phosphorus or floc settling rates were observed in any of the jar tests. It was, therefore, concluded that any benefit of solids-contact treatment would have to be demonstrated by pilot plant tests.

Anticipating that polyelectrolyte flocculation aids would improve chemical treatment floc settling rates, jar tests were conducted to quantify relative effects. Preliminary screening tests indicated that moderately anionic polyelectrolytes were effective flocculation aids for the three chemicals studied (alum, $FeCl_3$ and lime). Atlasep 2A2, a product of Atlas Chemical Industries, was used in subsequent jar test studies.

Figure 9 shows that lime treatment floc settling rates were increased three fold with a dosage of 0.75 mg/ℓ polyelectrolyte. It is also apparent that this polyelectrolyte dosage affected removal of substantial turbidity causing material not removed without polyelectrolyte.

Figure 10 shows the effect of polyelectrolyte on $FeCl_3$ treatment floc settling rates and supernatant turbidity. An optimum dosage of polyelectrolyte exists between 0.25 and 0.50 mg/ℓ. The significant increase in floc settling rates would superficially appear to warrant the use of polyelectrolyte with $FeCl_3$ treatment.

FIGURE 9: LIME TREATMENT: EFFECT OF POLYELECTROLYTE

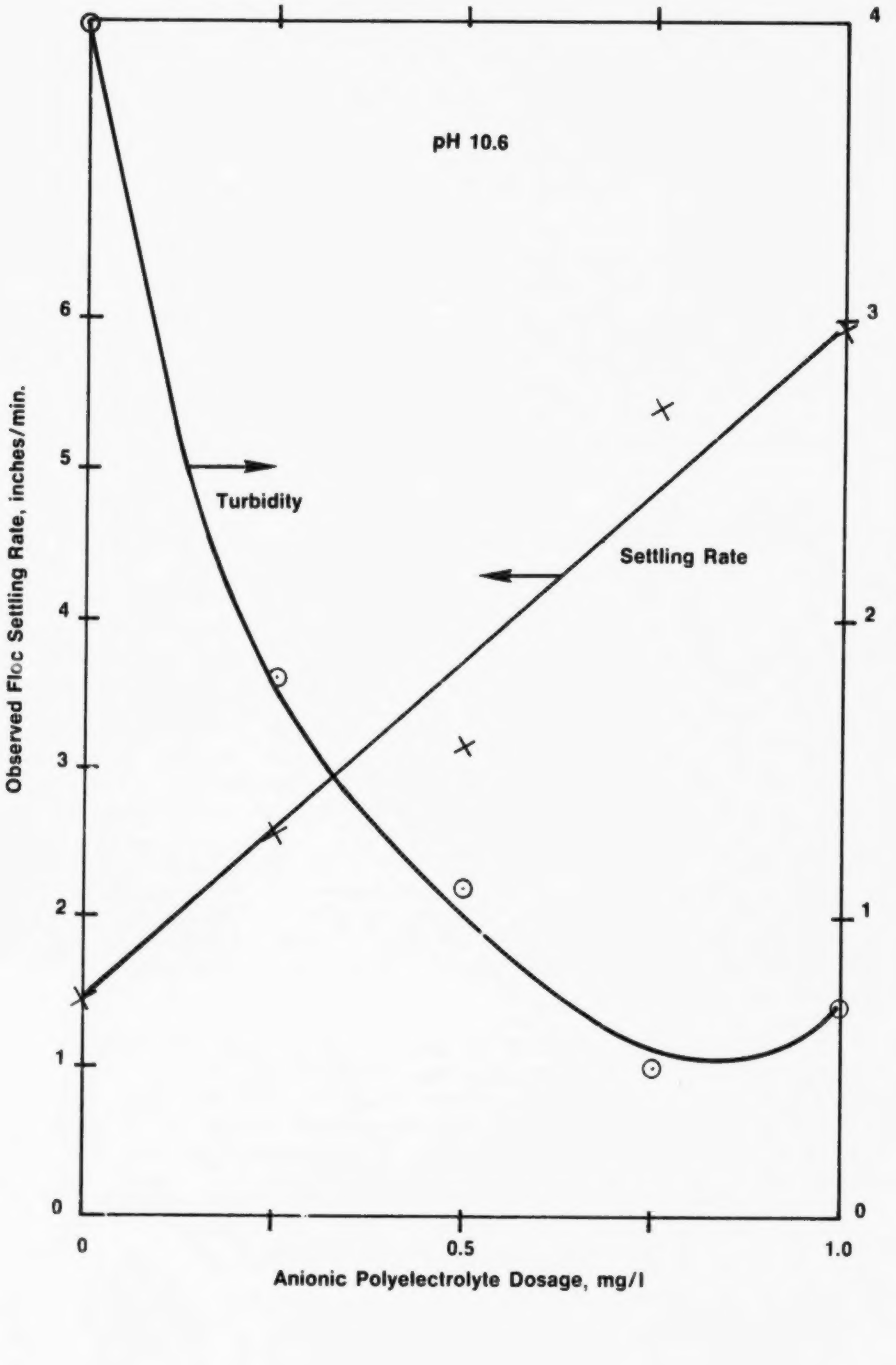
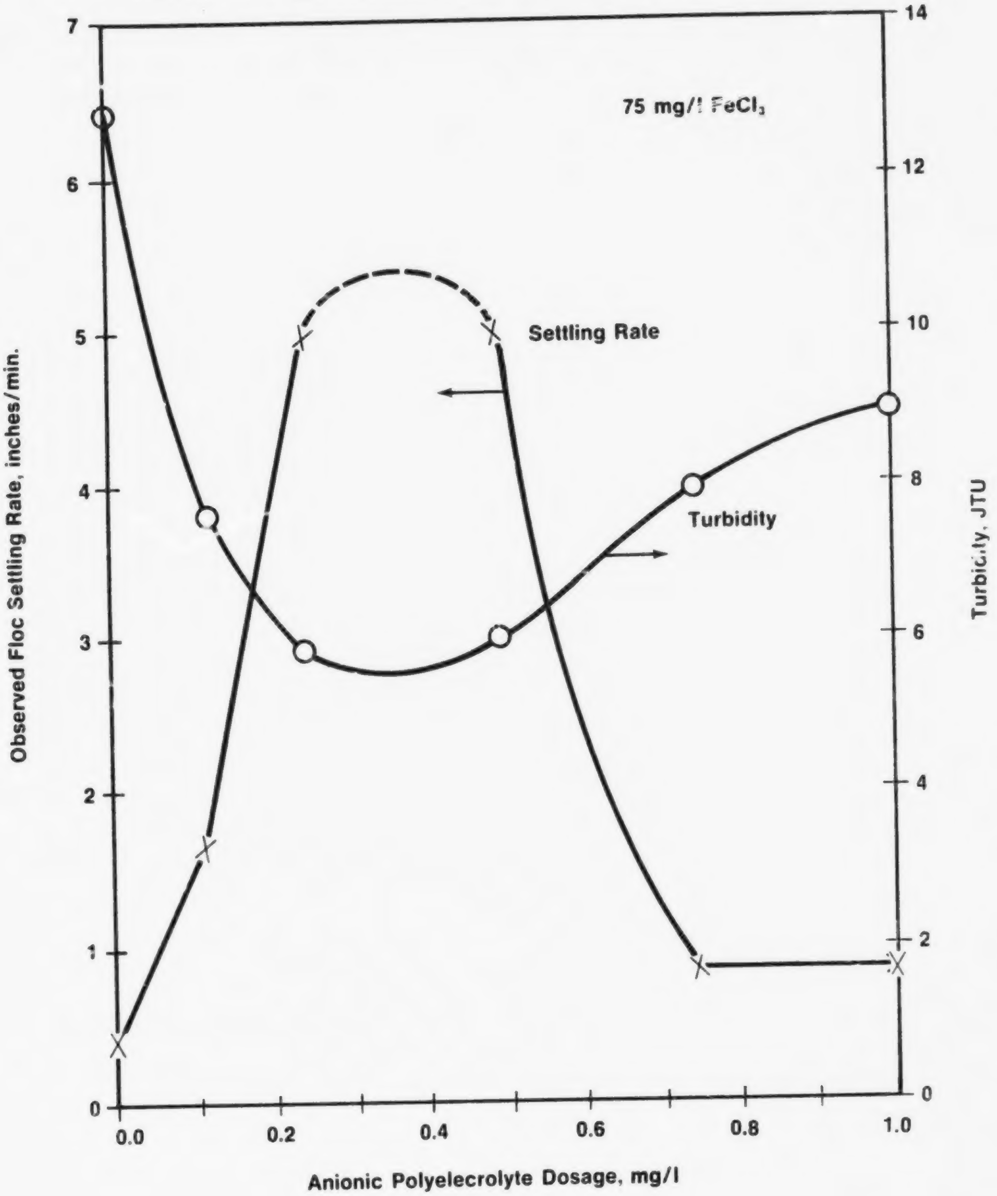


FIGURE 10: $FeCl_3$ TREATMENT: EFFECT OF POLYELECTROLYTE



POWDERED CARBON TREATMENT

Equilibrium adsorption isotherm tests were used to simulate removal of soluble organics by physical adsorption on powdered carbon. The TOC analyzer was not available when the laboratory adsorption tests were conducted so soluble COD was used as a measure of organic concentration. Total COD measurements of powdered carbon suspensions in tap water indicated that 1 mg/ℓ of powdered carbon (Aqua Nuchar A, a product of Westvaco Corporation) exerted a COD of about 1 mg/ℓ. Therefore, all samples, including blanks, were filtered through 0.45 micron membrane filters prior to COD determinations and the results reported as soluble COD (SCOD).

Brief screening tests failed to show any superior adsorption characteristics for four different commercial powdered activated carbons. The decision was made to use Aqua Nuchar A based on its relatively low cost and the fact that it had been used in several other relevant studies[14,15,16,17,23].

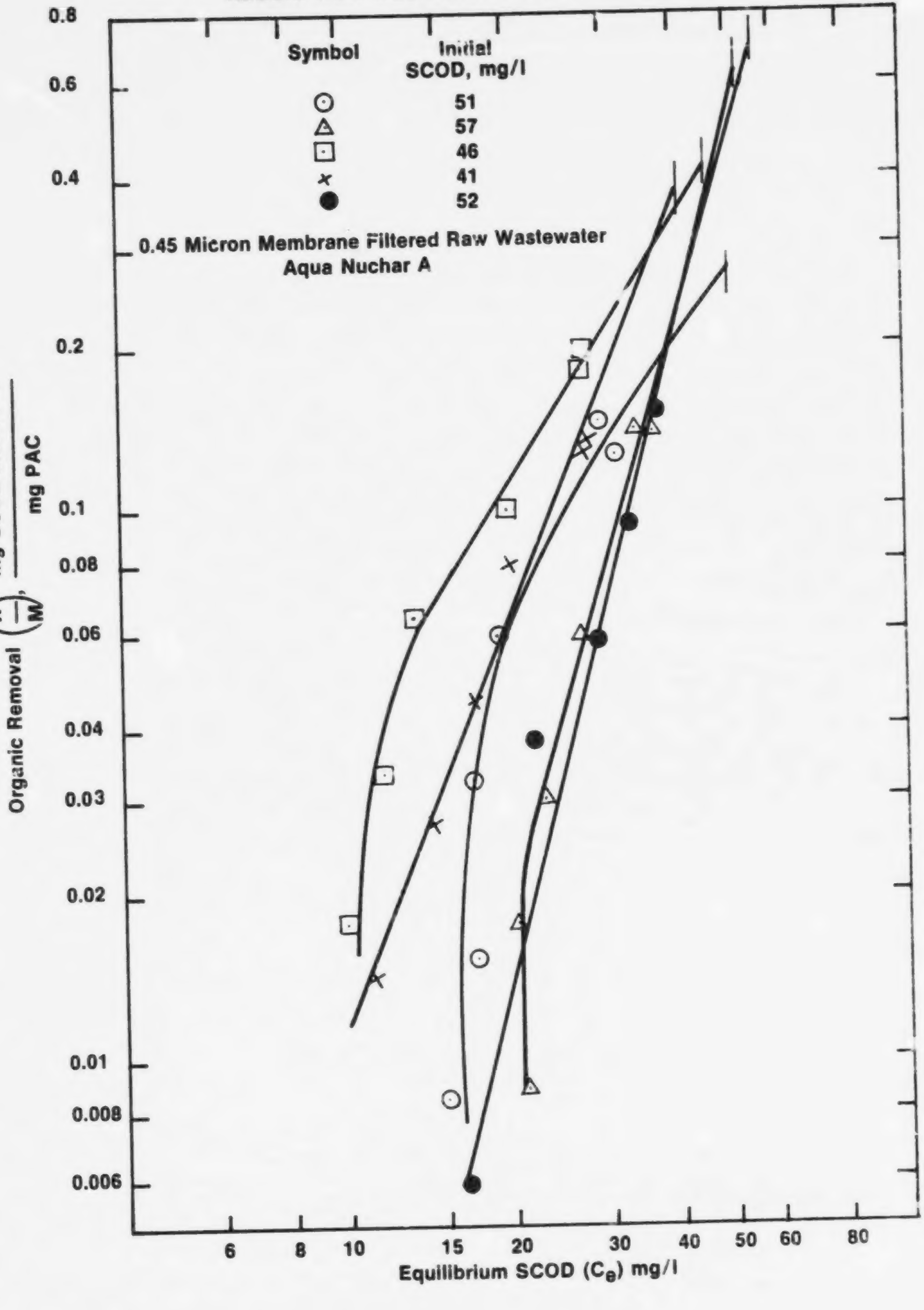
Figure 11 shows results of several equilibrium tests on raw wastewater. It is readily apparent that most of the equilibrium results cannot be modeled by the often used Freundlich equation. The trend of several curves toward a vertical asymptote at low values of equilibrium SCOD might be due to the existance of an unadsorbable SCOD fraction.

Equilibrium tests were conducted to determine if different chemical pretreatments affected the adsorption of SCOD on powdered carbon. Raw wastewater samples were coagulated-flocculated and settled using:

1. 150 mg/ℓ of $FeCl_3$
2. No chemical
3. Lime ($Ca(OH)_2$) to pH 12.0 ± 0.2

Supernatants from the settled samples were filtered through 0.45 micron membrane filters and equilibrium adsorption isotherm tests conducted using Aqua Nuchar A.

FIGURE 11: POWDERED CARBON TREATMENT: ADSORPTION EQUILIBRIUM ISOTHERM TESTS



Results of these tests are presented in Figures 12 and 13. The pretreatment did not significantly affect the equilibrium test results. It was, therefore, assumed that results of the pilot plant adsorption system would not be significantly influenced by the chemical used in the chemical treatment step.

Jar tests were conducted to determine the effect of carbon concentration and polyelectrolyte flocculation aids on gravity sedimentation. Jar test data in Figure 14 show that increasing carbon dosages improved supernatant clarity. For a dosage of 10,000 mg/ℓ, or 1 percent suspension. essentially all PAC was settled. This observation tends to substantiate the self-flocculating character of PAC suspensions indicated by other investigators[15,18].

Jar test screening of several types of polyelectrolytes indicated that anionic polyelectrolytes were most effective in increasing powdered carbon suspension settling rates. About 1 mg/ℓ of these flocculation aids resulted in about a threefold increase in floc settling rates to 1.5 inch/minute.

FIGURE 12: POWDERED CARBON TREATMENT:
EFFECT OF PRETREATMENT

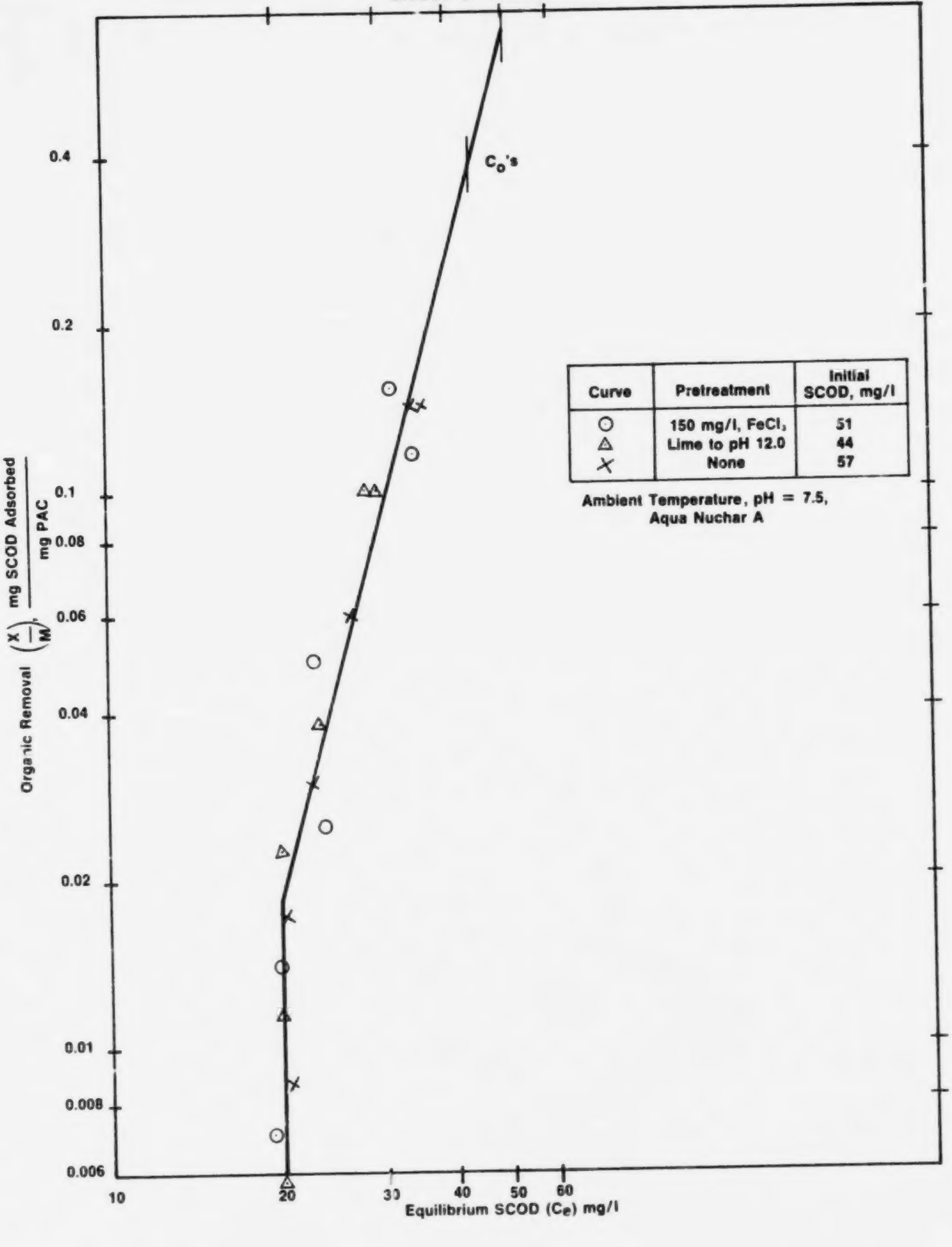


FIGURE 13: POWDERED CARBON TREATMENT: EFFECT OF PRETREATMENT

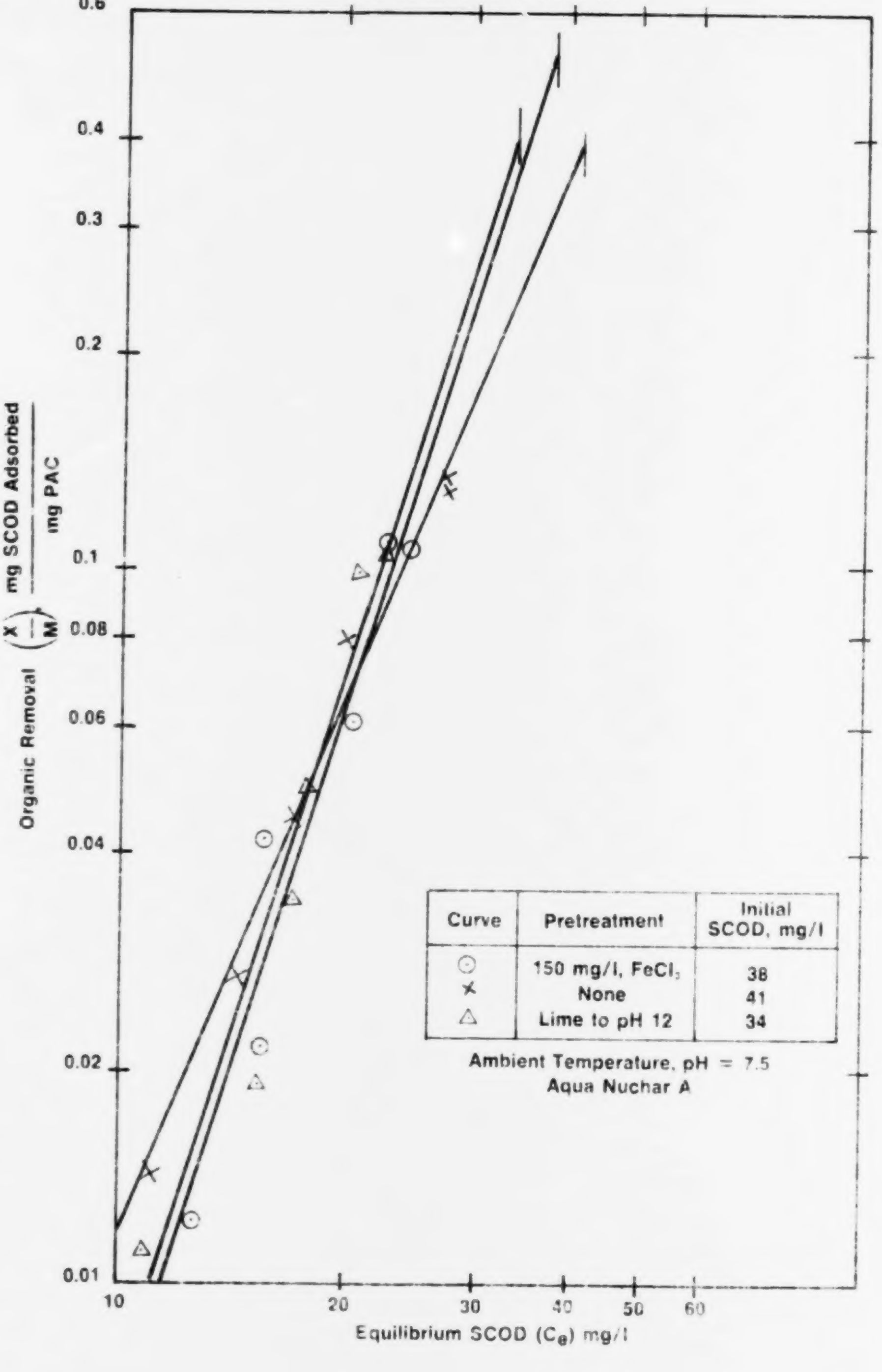
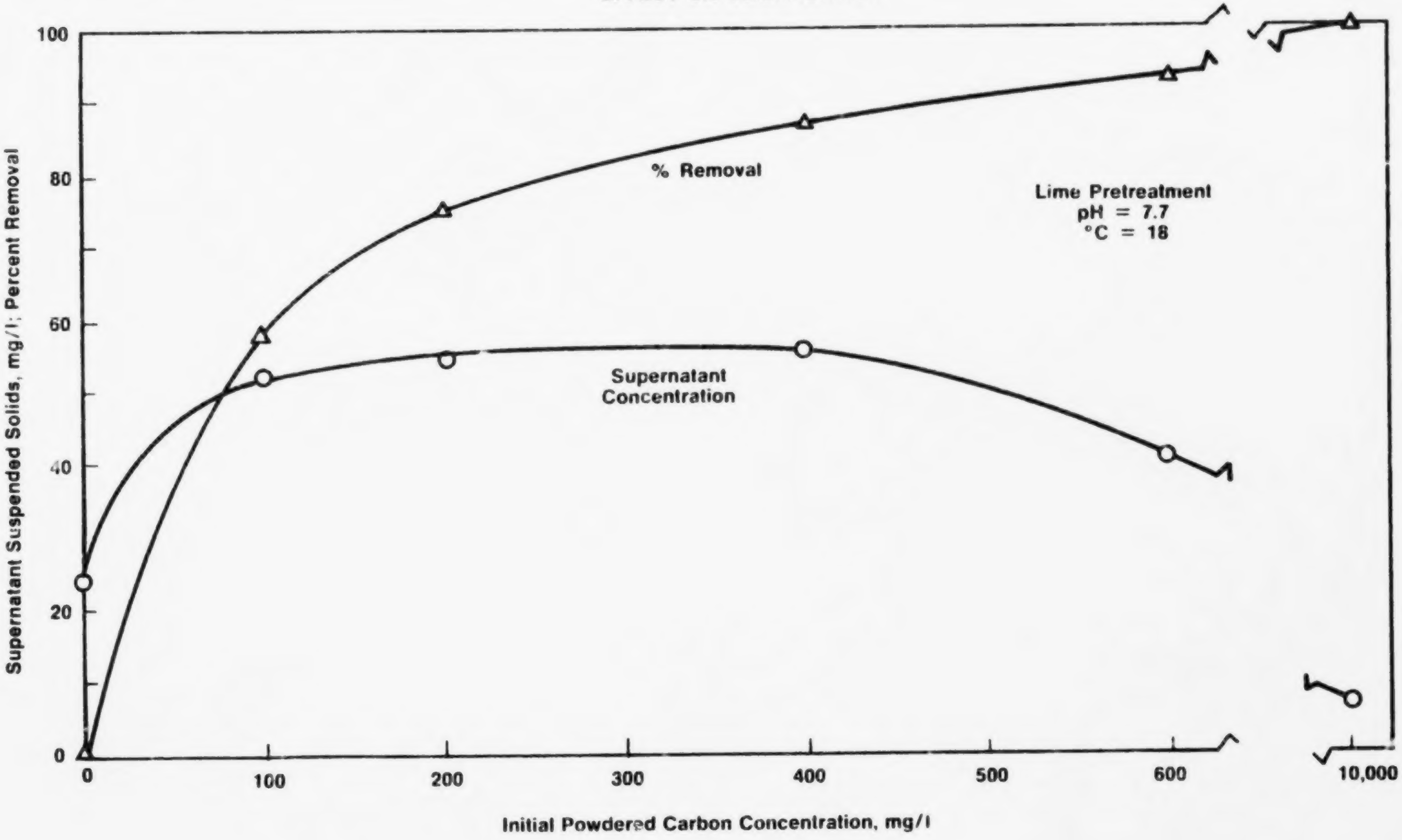


FIGURE 14: POWDERED CARBON TREATMENT: CONCENTRATION EFFECT ON CLARIFICATION



SECTION V

PILOT PLANT DESCRIPTION

RAW WASTEWATER FLOW

The pilot plant influent was coarse screened and comminuted Salt Lake City raw wastewater. About 350 gpm was diverted from the main Salt Lake City pump station discharge header to the pilot plant raw wastewater sump. Flow not used was returned to the pump station wet well by gravity. Some accumulation of very coarse grit occurred in the sump. The pilot plant raw wastewater pump was sized for a maximum flow of 150 gpm. The pump discharged to the chemical treatment unit via a magnetic flow meter-controller. Plant flow was automatically controlled at any constant flow rate equal to or less than 150 gpm. This flow was continuously recorded and totalized.

CHEMICAL TREATMENT UNIT

Chemical treatment and clarification of raw wastewater was accomplished in a 12 ft diameter Eimco Reactor-Clarifier solids-contact treatment unit, manufactured by Eimco Processing Machinery Division, Envirotech Corporation, Salt Lake City, Utah.

Figure 15 shows a schematic of this unit. Raw wastewater entered the unit via the influent pipe A to a draft tube E. Inorganic coagulants were added to the raw wastewater at B. Lime was added at point C. A slow speed turbine F pumped raw wastewater, chemicals, previously formed chemical precipitates and wastewater solids up the draft tube and into the reaction zone. The pumping turbine capacity was estimated to be about 5-8 times the average hydraulic throughput. Polyelectrolyte flocculation aids were introduced into the reaction zone at point D. Reaction zone slurry flowed under the reaction zone cone Q, into an increasing area upflow clarification zone. Settled solids were moved to the center of the unit by sludge rake, H, and either drawn up the draft tube or removed through the thickening cone I.

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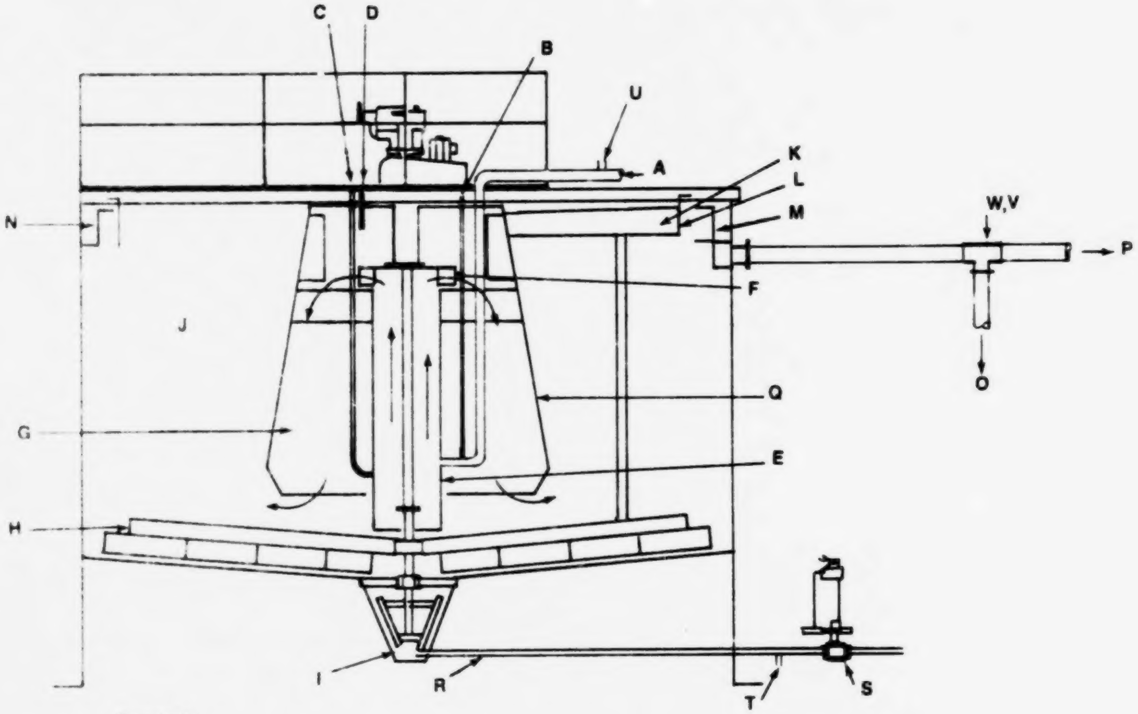
FIGURE 15: EIMCO REACTOR-CLARIFIER, SOLIDS-CONTACT CHEMICAL TREATMENT UNIT



Symbols:

A. Influent Pipe
B. Inorganic Coagulant Feed
C. Lime Feed
D. Polymer Feed
E. Draft Tube
F. Pumping Turbine
G. Sample Point, Reaction Zone
H. Sludge Rake
I. Sludge Thickening Cone
J. Expanding Area Clarification Zone
K. Liquid Surface Skimmer
L. Float Baffle
M. Peripherial V-Notched Launder
N. Acid Feed Point
O. Effluent to First Carbon Contactor
P. Effluent to Second Carbon Contactor or Drain
Q. Reaction Zone Cone
R. Sludge Blowdown Line
S. Timer Controlled Sludge Valve
T. Process Water Blowback Valve
U. Vacuum Breaker
V. Effluent Sample Point
W. Flow Splitter Box

Partially thickened sludge was automatically removed from the treatment unit by gravity flow through line R and a timer controlled sludge valve S. An automatic water backflush of line R was provided at valve T. Chemical sludge blowdown was routed to a 2000 gallon (8 ft. diameter) holding tank which was equipped with a turbine mixer.

Since the chemical treatment unit received raw wastewater, provision was made for clarification zone surface skimming by mechanism K. Floatable materials were kept from passing over the peripherial effluent launder M (a V-notched weir) by baffle L. Effluent neutralization was accomplished by addition of concentrated sulfuric acid to the effluent launder at point N. Clarified effluent exited the unit at splitter box W which divided flow to one or two carbon contactors, or to waste. A sample of clarifier effluent was continuously withdrawn at point V and conveyed to a composite sampling and process turbidity monitoring system. The approach to operation of the solids-contact unit used is presented in Appendix C.

CHEMICAL STORAGE AND HANDLING FACILITIES

Liquid alum or ferric chloride solutions were obtained in bulk quantities and stored in one of two 560 gallon (4 1/4 ft diameter) rubber lined tanks. Solution was removed from either tank at a point about 4 inches above the center of the bottom and fed by a positive displacement metering pump through PVC piping to point B in the chemical treatment unit influent line.

Concentrated sulfuric acid was obtained in bulk quantities and stored in a 1000 gallon covered mild steel tank. The acid was fed by a positive displacement metering pump through black iron piping to point N in the chemical treatment unit. An automatic acid neutralization system consisted of a process pH meter, recorder and controller. The pH probe was located in the chemical treatment unit effluent splitter box. The recorder had a low pH set point at which the acid feed pump was turned off. As the pH increases above the set point, the acid feed pump was turned on. Manual trim of the acid pump rate allowed minimizing frequent OFF-ON pump action.

High purity hydrated lime ($Ca(OH)_2$) was obtained in 50 pound bags. Dry lime was continuously slurried with tap water and pumped through plastic tubing by a positive displacement pump to point C (Figure 15) in the chemical treatment unit draft tube. The lime slurry pump was operated at a constant rate sufficient to prevent deposition in the feed lines. Lime dosage was varied by manually adjusting the dry lime feeder rate.

Two 560 gallon (4 1/4 ft diameter) polymer make-up tanks were provided for dissolving dry polymer into pilot plant effluent or tap water and storing polymer solution during use. High speed mixers (1750 rpm) were used to aid the solution of polymers and provide intermittent mixing of tank contents. The polymer solution was fed by positive displacement metering pumps through indicating flow meters to any of four treatment units (three solids-contact units and the granular media filter).

PAC CONTACTORS

PAC treatment of chemically clarified raw wastewater was accomplished in one or two 10 ft diameter solids-contact treatment units. These units were identical in detail to the chemical treatment unit (Figure 15) with the exception that they did not have surface skimmers and that submerged orifice launders were used. The units were piped to be operated either in parallel, as single-stage carbon contactors, or in series, as a two-stage counter-current carbon contacting system.

Partially thickened spent PAC slurry was automatically removed from each contactor by gravity flow.

Virgin PAC was made up in a 840 gallon (5 ft diameter) rubber lined tank containing a mixer. When available, pilot plant effluent was used, otherwise tap water was used to wet and slurry the PAC. PAC was obtained in 35 pound bags and manually poured, very slowly, into the make-up tank half filled with water. The make-up tank had a hinged plywood cover to prevent gross dusting problems during initial mixing of the PAC. Normally, 12 to 24 hours mixing was allowed for thorough wetting before transferring the PAC slurry to a second 840 gallon rubber lined feed tank.

Virgin PAC slurry was fed to the influent line of the carbon contactors by a positive displacement (ball check) pump. When the two carbon-contactors were operated in series, virgin PAC slurry was fed only to the second-stage unit. Underflow from this unit was intermittently pumped by a peristaltic pump to the influent line of the first-stage unit. Spent PAC exited the first-stage unit via blowdown to a thickener. No attempt was made to identically match the PAC mass transfer rate between units with the virgin PAC mass feed rate. Rather, a PAC transfer rate was maintained at that required to enable stable operation of the second-stage solids-contact carbon contactor. For this reason, it may have taken 2-4 days of plant operation to reach steady-state PAC mass flow. The approach to operation of the solids contact treatment units used is presented in Appendix C.

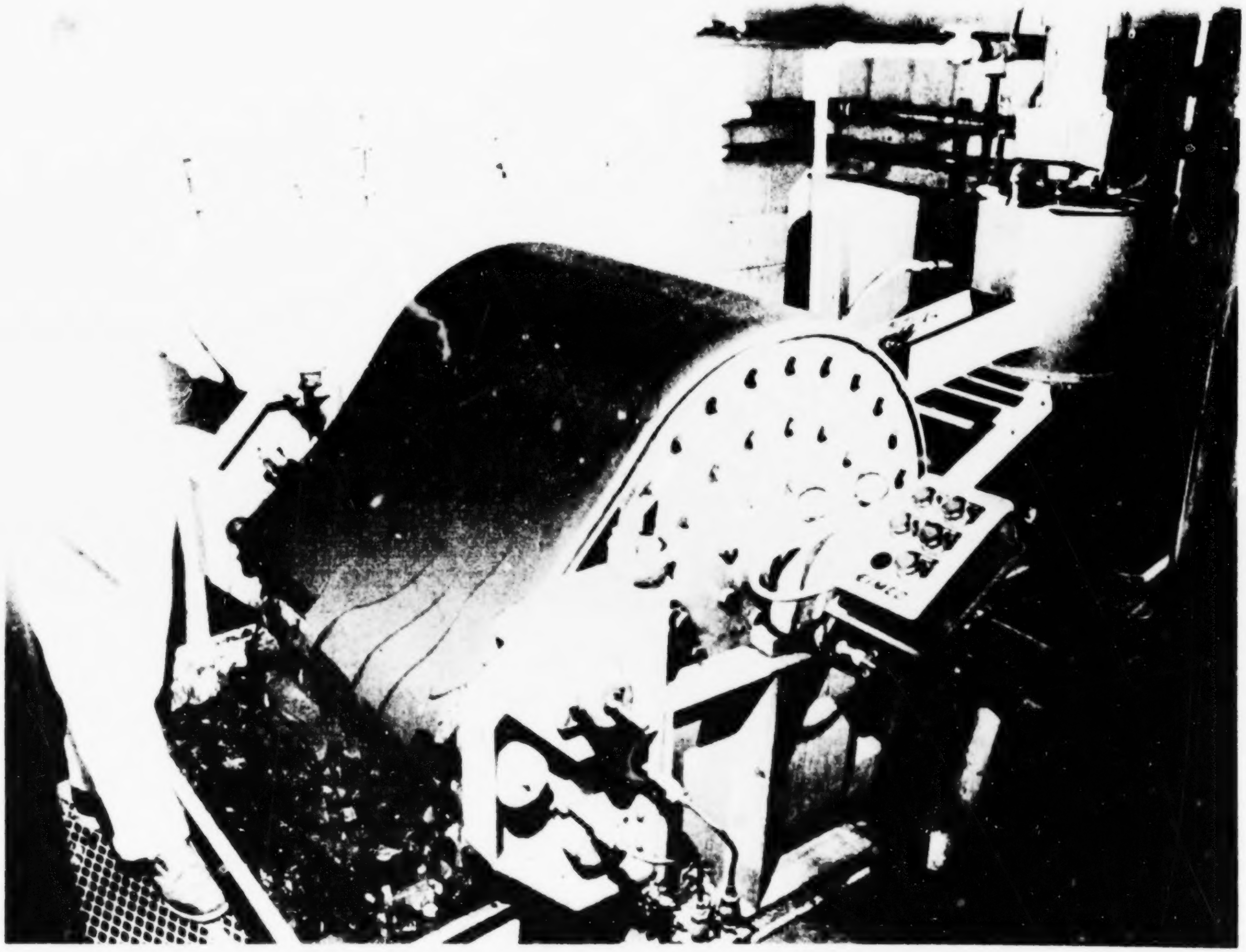
POWDERED CARBON REGENERATION SYSTEM

The regeneration system consisted of four major unit operations: carbon sludge concentration by gravity thickening, vacuum dewatering, regeneration in a fluidized bed furnace and off-gas carbon capture by wet venturi scrubbing. This system is shown as part of Figure 1.

Spent carbon was removed from the carbon contactors by gravity flow to a five (5) ft diameter gravity thickener. Thickener underflow was transferred to a carbon inventory and sampling tank (about 500 gallons). A small peristaltic type pump was used for this transfer operation. The slurry volume of thickened carbon was measured. Then, after thorough mixing, a representative sample was obtained for laboratory tests. The inventoried, thickened carbon was then transferred to a 1700 gallon conical storage tank. To prevent bridging in the conical storage tank, carbon was recirculated, externally, at a rate of about 5 gpm from the bottom outlet to the top of the tank.

Vacuum filter dewatering was accomplished with a three (3) ft diameter by three (3) ft face EimcoMet Eimco-Belt filter manufactured by Eimco Processing Machinery Division, Envirotech Corporation, Salt Lake City, Utah. The vacuum filter station consisted of the filter, flocculation tank, vacuum pump, filtrate pump and receiver all mounted on a platform. Figure 16 shows the vacuum filter in operation. All components of the filter which

FIGURE 16: VACUUM FILTRATION OF SPENT POWDER CARBON



contacted carbon slurry were made of corrosion resistant thermoplastic. A rubber lined chemical flocculation tank had two mixing compartments to achieve chemical dispersion and floc growth. A 55 gallon drum, high speed mixer and variable speed positive displacement diaphragm chemical feed pump comprised the conditioning chemical make-up and feeding facility.

The filter cake discharged into a hopper which had a capacity of about 300 lb of wet filter cake. A ribbon screw advanced the cake along the bottom of the hopper to the throat of the feed pump. The ribbon screw reconstituted the filter cake for form a paste which the pump readily transported to the fluidized bed furnace.

A schematic of the furnace is shown in Figure 17 and a photograph of the unit in Figure 18. The outer dimensions were 5 1/4 ft diameter by 14 ft high. The outer steel shell was lined with suitable fire and insulating brick and castable refractory to provide and following approximate internal dimensions:

1. 3 ft diameter by 3 ft high fire box
2. 3 ft diameter by 3 ft high sand bed chamber
3. 3 1/4 ft diameter by 6 ft high freeboard chamber

Thermocouples and pressure taps (P/I) were located as shown in Figure 17. Stainless steel tuyeres (nozzles) were provided to retain the sand and provide distribution of fire box gases into the sand bed.

The fuel used was natural gas. Combustion air was provided by a 200 SCFM capacity blower. Injection of natural gas directly into the fluidized bed was provided by six nozzles, equally spaced around the circumference. These nozzles entered the bed just above the tuyeres.

Natural gas was fired in the burner chamber with about 150 percent (by volume) excess air to keep the fire box temperature below about 2000 °F. This limited temperature was required to protect the refractory and stainless steel tuyeres. Hot gas flow from the fire box at a superficial velocity of about one (1) ft per second was required to maintain the 15 x 30 U.S. Mesh sand bed in a fluidized state. Prior to feeding carbon, the oxygen content of the furnace off-gas was trimmed with bed injection gas feed. Dewatered carbon was automatically fed at a rate which maintained the bed temperature at a preset value.

FIGURE 17: FLUIDIZED BED REGENERATION FURNACE

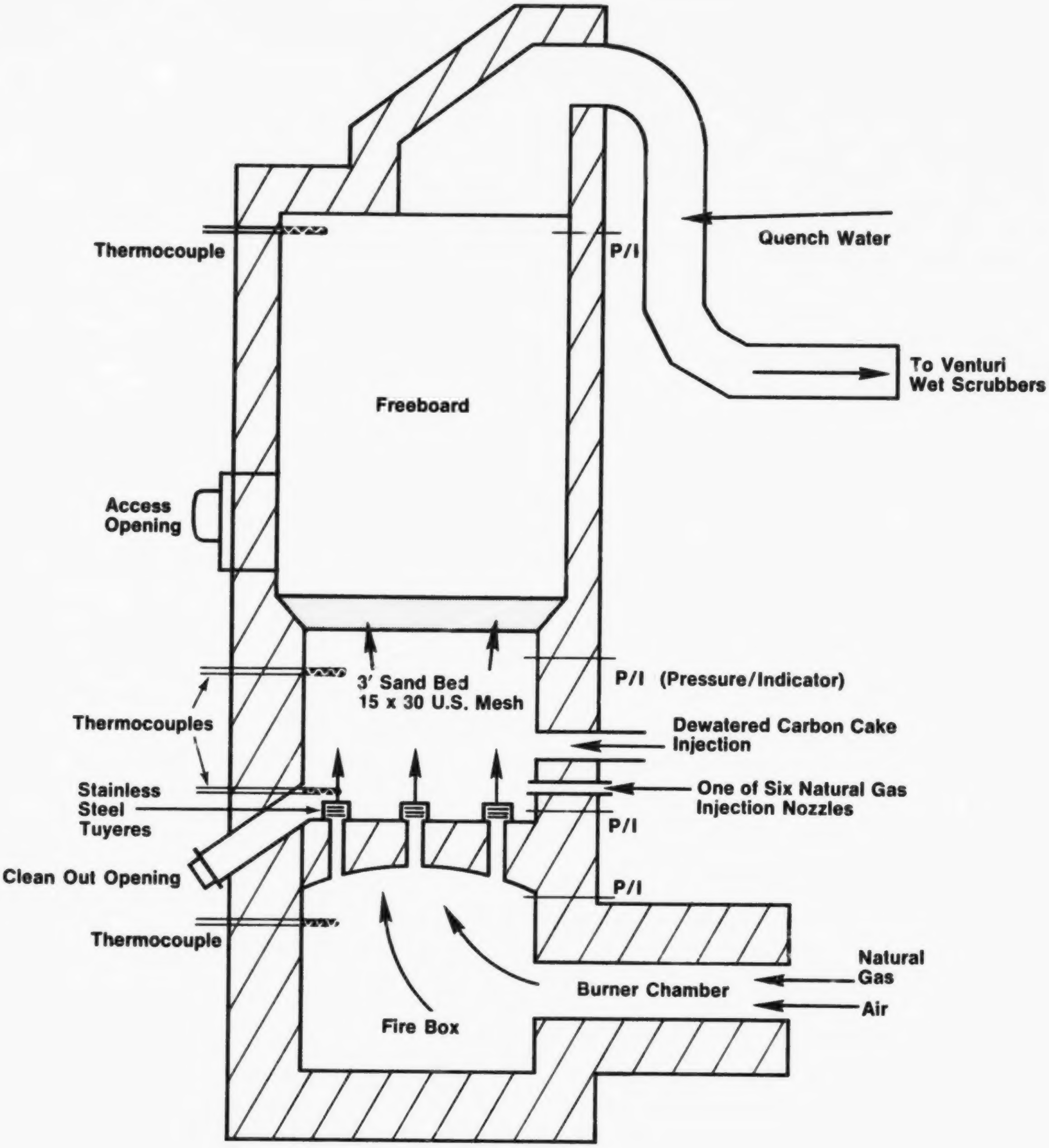
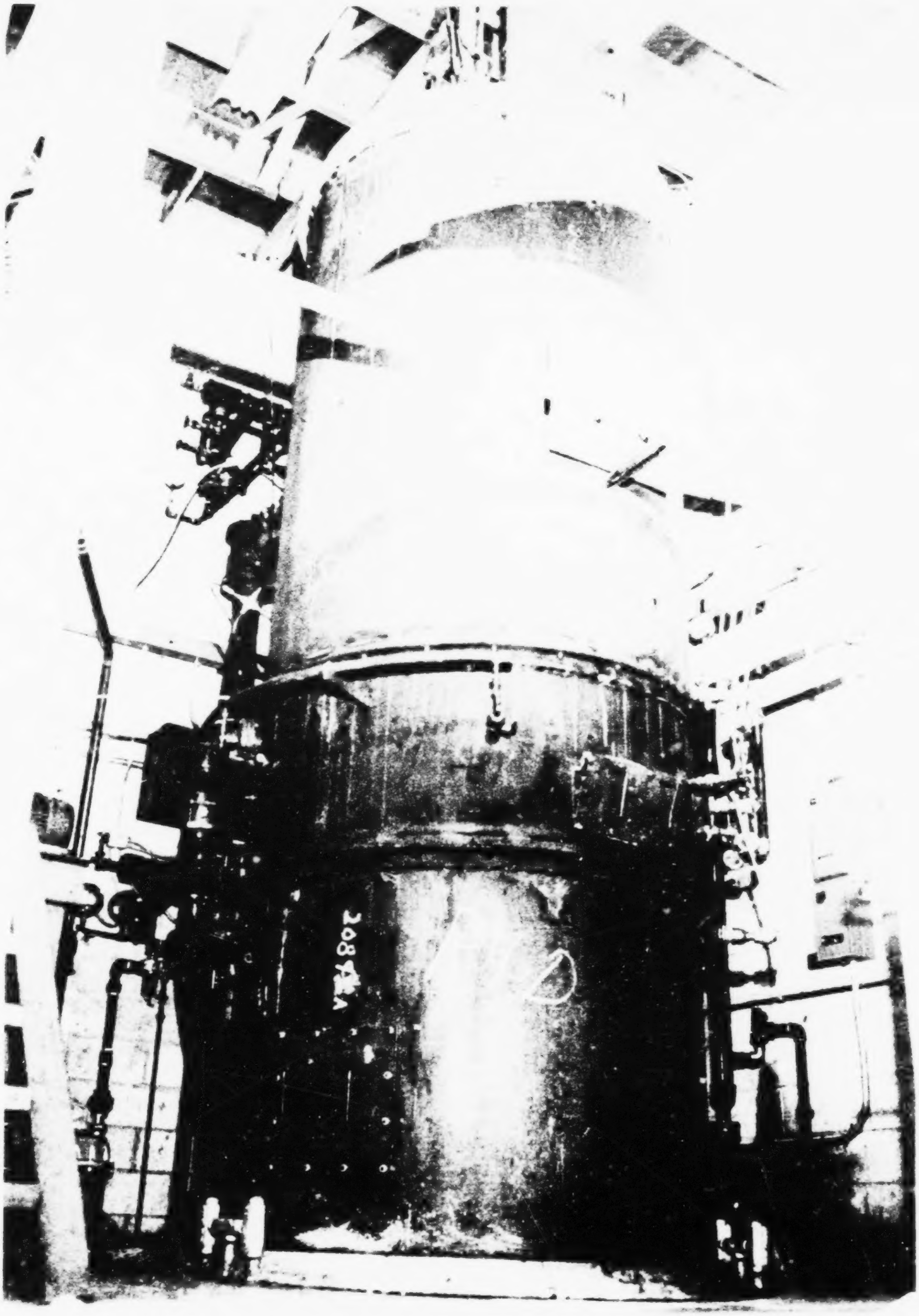

Thermocouple
P/I
Quench Water
To Venturi Wet Scrubbers
Freeboard
Access Opening
3' Sand Bed 15 x 30 U.S. Mesh
P/I (Pressure/Indicator)
Thermocouples
Dewatered Carbon Cake Injection
Stainless Steel Tuyeres
One of Six Natural Gas Injection Nozzles
P/I
Clean Out Opening
P/I
Thermocouple
Natural Gas
Burner Chamber
Air
Fire Box

FIGURE 18: PAC FLUIDIZED BED REGENERATION FURNACE



The hot furnace off-gasses, containing the regenerated carbon, were quenched to about 200°F by the addition of about 15 gpm of recycled scrubber water to the furnace outlet duct. The cooled gases were then passed through a 6 inch and then a 4 inch diameter wet venturi scrubbers, in series to capture essentially all of the regenerated carbon. Scrubber water flow was about 30 and 15 gpm to the 6 and 4 inch venturies respectively. Scrubber water underflow was discharged to a 1700 gallon conical decant tank which provided some separation of carbon. Decant tank overflow was pumped through a heat exchanger, prior to reuse as quenching and scrubber water. Pilot plant effluent was used for heat exchanger water. The fluidized bed furnace and venturi scrubbing system were highly instrumented for safety and ease of operation.

Carbon regeneration runs were conducted on a batch basis. All the stored, thickened spent carbon was run through the vacuum filter and regeneration furnace system at one time. Spent carbon not reaching the furnace was accounted for by collecting all vacuum filter filtrate, spillage and all washings containing spent carbon. An isolated regeneration room floor drain system with an intercepting sump facilitated this operation.

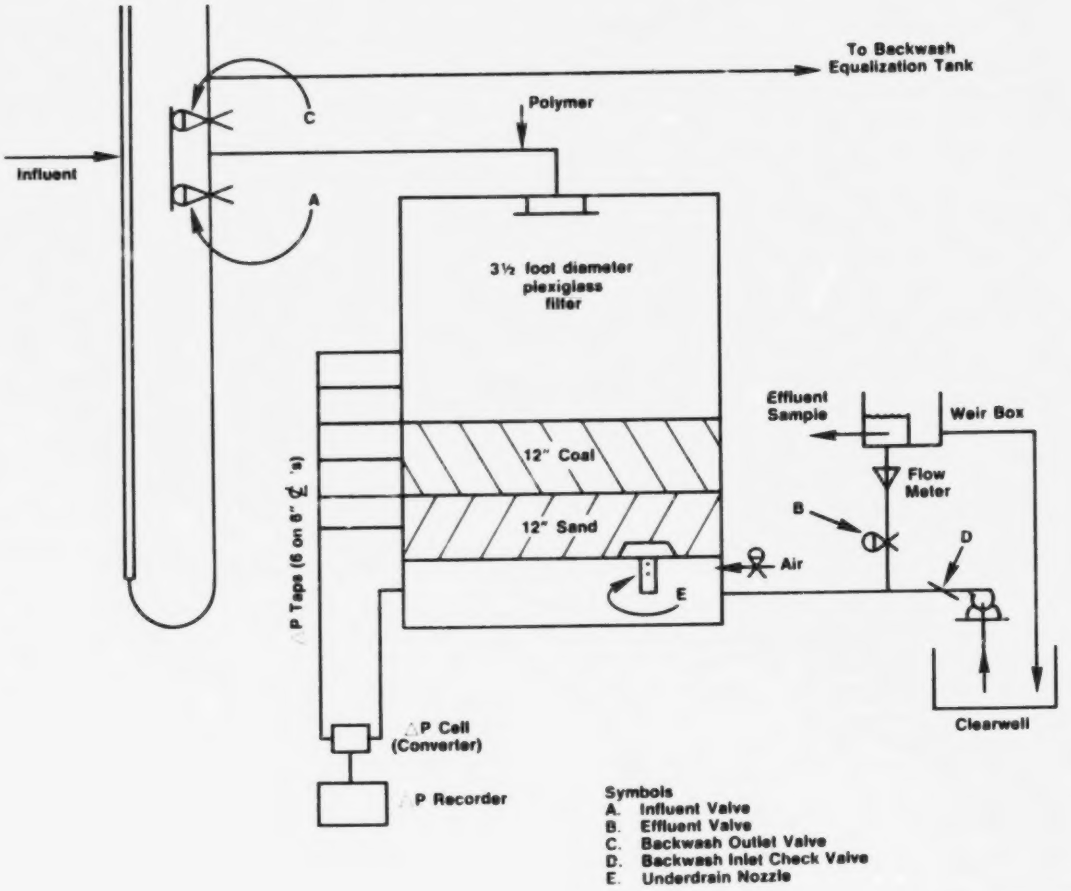
GRANULAR MEDIA FILTRATION STATION

Effluent from one of the carbon contactors was applied to the granular media filter. The filter was a 3.5 ft diameter by 6 ft high clear plastic unit of conventional downflow gravity design with a coal-sand filter bed. Backwashing of the filter bed was accomplished by an automated air-scour hydraulic method.

The sand layer of the dual media filter bed was 12 inches of 20 x 30 U.S. Mesh filter sand. The coal layer was about 12 inches of specially sized anthracite coal. The coal effective size was 1.1 mm and the uniformity coefficient was less than 1.4, after installation and removal of two inches of fines from the top of the backwashed and settled filter bed.

Figure 19 is a schematic of the granular media filtration station. In the filtering mode, influent flowed through a U-tube to the filter inlet. The dual-media filter bed rested directly on a false bottom. Flow was collected in the filter outlet plenum and routed to an effluent weir

FIGURE 19: GRANULAR MEDIA FILTRATION STATION


To Backwash
Equalization Tank
Polymer
C
Influent
A
3½ foot diameter
plexiglass
filter
Effluent
Sample
Weir Box
12" Coal
Flow
Meter
B
12" Sand
D
Air
E
ΔP Taps (6 on 6" ℄ 's)
Clearwell
ΔP Cell
(Converter)
ΔP Recorder
Symbols
A. Influent Valve
B. Effluent Valve
C. Backwash Outlet Valve
D. Backwash Inlet Check Valve
E. Underdrain Nozzle

box located at an elevation which insured that liquid in the filter bed was never at less than atmospheric pressure. Filter flow rate was indicated by an orifice-plate bypass flow meter. A sample of filter effluent was continuously pumped from the effluent weir box to a composite sampler and process turbidimeter. Clarifier wastewater flowed from the effluent weir box to a 5000 gallon (12 ft diameter) clearwell which provided treated water storage for plant use and filter backwash.

Two columns of pressure taps (90° apart) were located at 6 inch intervals up the side of the filter housing. Pressure differentials (ΔP) between the outlet plenum and each pressure tap were automatically measured and recorded once each hour for a duration of 8-10 minutes. The filter was automatically backwashed when the total filter bed headloss (including underdrain) reached a preset value which was routinely set at 7 ft of water. The operational sequence during backwashing is presented in Appendix D.

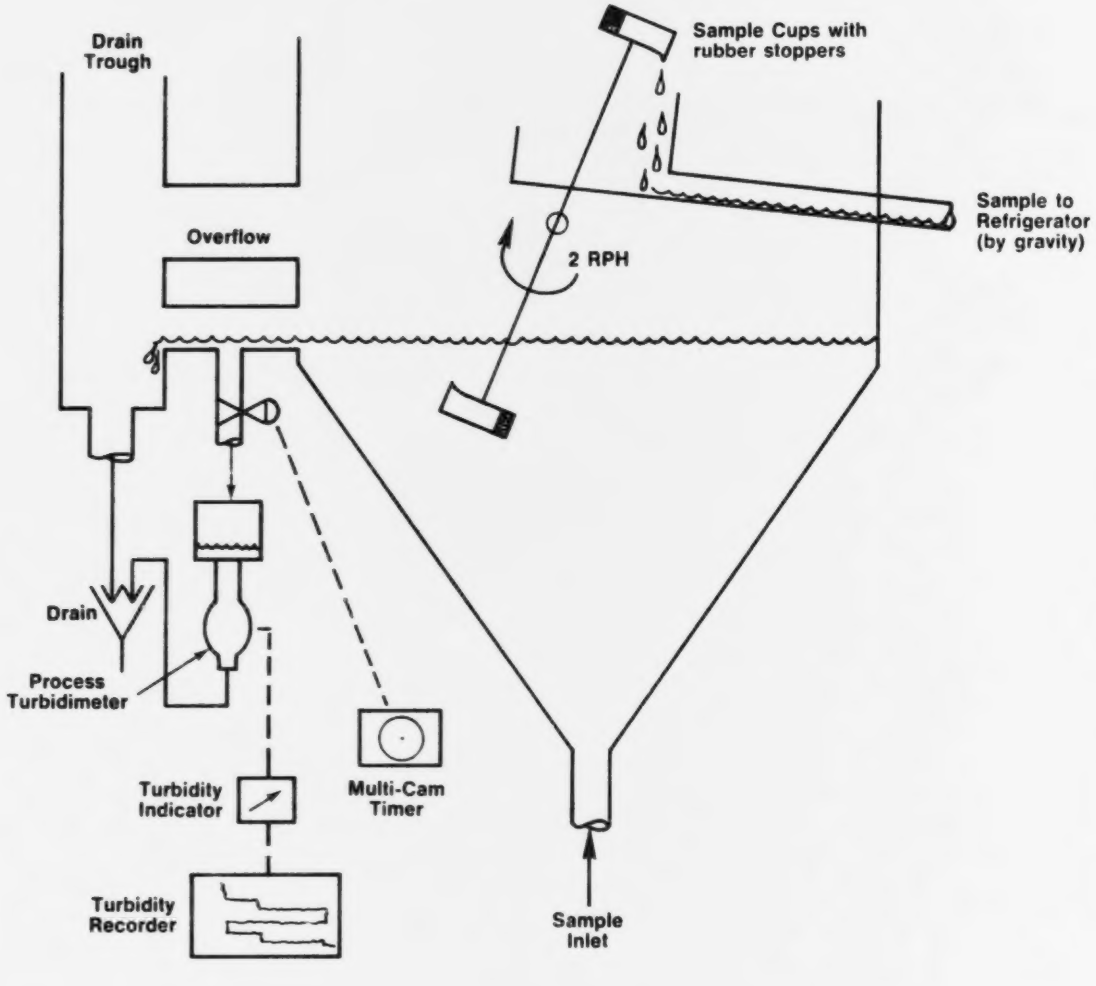
Backwash water was collected in a 3600 gallon (10 ft diameter) equilization tank which was equipped with a low speed mixing turbine. Equilization tank contents were pumped to the reaction zone of the first-stage carbon-contactor at a rate of about 5 gpm.

COMPOSITE SAMPLER AND PROCESS TURBIDIMETER

An automatic sample compositor and turbidity measuring and recording system was developed for this project (see Figure 20). All process stream samples (five) were piped to a central location. Each stream passed upward into a hopper shaped container, sized to maintain solids in suspension. Sample continuously overflowed, the hopper via a 1 inch diameter horizontal pipe to a common waste drain. Normally-closed solenoid valves were located in a vertical line just below the horizontal pipes. Outlets from all five solenoids emptied into a common trough which drained into a flow-through process turbidimeter. A multiple cam timer programmed each solenoid to be opened for 12-15 minute intervals once each hour. The raw wastewater sample was not monitored for turbidity, since continual plugging of the solenoid valve was experienced.

Twenty-four hour, time composite samples were obtained from the flow through the hopper container mentioned previously. Approximately 30 ml of samples were collected in plastic cups

FIGURE 20: COMPOSITE SAMPLING AND PROCESS TURBIDITY MONITORING DEVICE


Drain Trough
Overflow
Sample Cups with rubber stoppers
Sample to Refrigerator (by gravity)
2 RPH
Drain
Process Turbidimeter
Turbidity Indicator
Multi-Cam Timer
Turbidity Recorder
Sample Inlet

each 30 minutes. The plastic cup bottoms had rubber stoppers to facilitate cleaning. Samples were transferred from the sample collection trough to refrigerated storage vessels. Composited samples were refrigerated at about 34°F until removed for analysis.

LABORATORY FACILITIES

Analytical laboratory facilities were located at the pilot plant site. The laboratory was equipped to conduct all analyses reported except ammonia nitrogen, BOD_5 and sulfides. These analyses were made at the Eimco Corporation's Sanitary Engineering Research and Development laboratory. Sludge thickening and vacuum dewatering tests were also conducted at the Eimco laboratory.

SECTION VI

APPROACH TO PILOT PLANT OPERATION

Initially, the approach to the pilot plant test program consisted of choosing desired effluent water quality criteria for each unit operation. The plant was then operated to determine optimum conditions (e.g., minimum chemicals, maximum flows, etc.) which would result in the desired effluent quality. Emphasis was placed on predicting desirable plant operating conditions based on laboratory test results. For example, chemical coagulant dosage and polymer type and dosage were obtained from standard jar tests.

The above approach was followed during the first 3 months of plant operation with less than satisfactory success. In general, the plant was "over operated" by frequent changing of operation variables such as $FeCl_3$ and polyelectrolyte feed rates. Because of the hourly and daily variability of the raw wastewater, minimal available operator time and a six day per week plant operating schedule, meaningful process results were extremely difficult to obtain. Consequently, the following approach was adopted and adhered to for the remainder of the test program:

1. Based on previous operating experience and/or laboratory jar test results, operating conditions were established (chemical type and dosage, PAC dosage and operating mode).

2. The plant was brought on a stream at a nominal hydraulic loading for two to three days to establish near steady state conditions. The flow was then increased by 10-20 percent increments each day, until the effluent clarity consistantly exceeded about 25 JTU. This observation indicated that the maximum hydraulic loading had been reached. The flow was then reduced to 75-85 percent of the maximum established.

3. The plant was operated under the established operating criteria for one to three weeks during which time twenty-four hour composite samples were collected and analyzed and treatment efficiency established.

CHEMICAL MONITORING

Alum and $FeCl_3$: Each new batch of these chemical solutions were immediately sampled and the Al^{+3} or Fe^{+3} weight concentration and solution density determined. Typical solution strengths were 42 percent by weight $FeCl_3$ and 15 percent by weight Al_2O_3. The actual solution volumetric feed rates were measured four to ten times daily. In addition, daily volume inventory data were collected on each chemical feed tank, along with the total flow through the chemical treatment unit. The average dosage rate was established from these average daily values.

Lime: A lime availability was run on every second or third shipment of 100 fifty pound bags of high quality lime, $Ca(OH)_2$. Availabilities of about 90 percent were observed. The rate of lime feed was based on observed pH values of samples from the reaction zone of the chemical treatment unit. If the observed pH was more than 0.2 units above or below the specified operating level, the rate of dry lime feed was appropriately adjusted. This manual control of treatment pH was generally acceptable for the constant-flow plant operation conditions. However, on several occasions, bridging in the lime feed hopper and changes in the raw wastewater alkalinity and/or pH caused significant departures from desirable operating pH with noticeable treatment upset resulting. The actual dry lime weight feed rate was measured four to ten times per day and averaged. In addition, the total number of bags of lime used was recorded and an average lime feed rate computed using the total flow through the chemical treatment unit.

Acid: Concentrated acid (94 percent H_2SO_4 by weight) feed rate was determined by a daily inventory of the acid storage tank and the chemical treatment unit total daily flow.

Polymer: Polymer solutions of 0.025 percent to 0.050 percent were made up according to supplier recommendations. Monitoring of polymer feed rates were identical to the procedure used for alum and $FeCl_3$.

Powdered Activated Carbon: PAC feed slurry concentrations of 2.5 and 5.0 percent by weight were used. The 2.5 percent concentration was used for low plant flow and/or low PAC dosage conditions. The volumetric feed rate of PAC slurry was measured four to ten times per day. A daily inventory of the PAC feed tank contents was made along with a measurement of total flow passing through the carbon contactors. The average daily dosage was determined from these observations.

PILOT PLANT START-UP

Numerous start-up problems were encountered. Typical problems involved inadequate plumbing and electrical circuitry, underdesigned mixing equipment, and PAC slurry corrosion. The more important equipment operation problems will be discussed later.

With the many mechanical start-up problems, two months were required to achieve a acceptable level of operator training and plant operation. During this period, the plant was onstream no more than six days per week (it was shut down on Sundays). Operators were on duty only during about two-thirds of the time. This situation proved untenable in as much as frequent mechanical or process failures occurred during times that operators were not present, resulting in unsatisfactory plant performance. Also, after a day of plant down time, at least a day was required to re-establish steady state treatment conditions. Therefore, after the third month of operation, it was decided to increase the operating staff and keep the pilot plant on-stream continuously except during holidays.

SCHEDULE OF PLANT TEST

Because laboratory results indicated that chemical pretreatment did not significantly effect the adsorption of SCOD on PAC, and because of multiplicity of treatment variables were to be studied, the chemical treatment and PAC adsorption tests were independently scheduled. The following is a listing of the major variables studied.

A. Chemical treatment:

    1. Ferric chloride
        a. Overflow rate
        b. $FeCl_3$ dosage
        c. Need for polyelectrolyte
    2. Lime [$Ca(OH)_2$]
        a. Overflow rate
        b. Treatment pH
    3. Alum
        a. Overflow rate
        b. Need for polyelectrolyte

B. PAC contacting

    1. Two-stage counter-current contacting

        a. PAC dosage
        b. Need for polyelectrolyte
        c. Overflow rate

    2. Single-stage contacting

        a. PAC dosage
        b. Overflow rate

C. Granular Media Filtration

    1. Filtration rate

    2. Need for polyelectrolyte

SCHEDULE OF ANALYTICAL TESTS

The basic philosophy during this pilot plant study was that the following priority of functions be followed:

No. 1: Steady state plant operating conditions were well established.

No. 2: Representative samples of all process streams were collected.

No. 3: A spectrum of analytical tests on the composite samples of the various process streams were made.

If priority items 1 or 2 were not satisfied then item 3 was not undertaken.

The spectrum of analytical tests run on composite samples is presented in Table 1. In addition, suspended solids analysis were run on samples of the reaction zone slurry and sludge blowdown from each chemical and PAC solids-contact treatment units. These samples were 24-hour composites of grab samples taken at two hour intervals.

At weekly to bi-weekly intervals, a sample of 24-hour composited chemical-sewage sludge was collected and laboratory thickening and vacuum filter leaf tests conducted.

TABLE 1: SCHEDULED ANALYTICAL TESTS OF COMPOSITE SAMPLES

			PAC CONTACTOR		
Water Quality Parameter	Raw Waste-water	Chemical Treatment	First Stage	Second Stage	GM Filter
Turbidity	a	a	a	a	a
Suspended Solids	a	a	a	a	a
Total Phosphorus	a	a	-	-	a
Soluble Total Phosphorus	a	a	-	-	-
Total Organic Carbon	a	a	a	a	a
Soluble Total Organic Carbon	a	a	a	a	a
Chemical Oxygen Demand	a	a	-	-	a
Soluble Chemical Oxygen Demand	a	a	a	a	a
Biochemical Oxygen Demand	b	b	b	b	b
pH	a	a	a	a	a
Total Alkalinity	a	a	-	-	a
Total Hardness	a	a	-	-	a
Calcium Hardness	a	a	-	-	a
Sulfides	b	b	b	b	b
Ammonia - Nitrogen	b	b	b	b	b
Total Soluble Iron	c	c	c	c	c

---

a - Routine analysis on each 24-hour composite
b - No more frequently than once per week
c - Intermittent during $FeCl_3$ campaign

PLANT OPERATING DATA

Plant operators routinely monitored and recorded equipment and process data, 5 to 10 times per day, to assure that the plant was operating under specified conditions. All chemical and process flows were observed or measured for all the solids-contact units, as were pumping turbine speed, rake speed and torque, and blowdown conditions (time interval and duration).

Two liter samples of each solids-contact unit reaction zone slurry were collected every two to four hours and 5-minute settling tests conducted. The location of the slurry pool in the clarification zone of each unit was also observed every two to four hours by drawing samples from the three sample taps. The highest sample tap at which copious amounts of sludge appeared was recorded.

Turbidity, pH and temperature measurements were made of grab samples of all process streams taken at 2 to 6 hour intervals.

The comprehensive plant operating data collection was considered desirable to eliminate potential treatment upsets due to malfunctioning of equipment and variable raw wastewater characteristics. Had longer plant test compaigns been possible, less surveillance would have sufficed.

SECTION VII

PILOT PLANT RESULTS

Pilot plant results will be presented and discussed in the following separate sections.

1. Chemical treatment step:

   The purpose of this treatment step was to provide the majority of phosphorus and suspended solids removal. Other physical-chemical treatment schemes and some chemical-biological treatment schemes could include this chemical treatment step. Thus, chemical treatment results are presented separately.

2. Powdered activated carbon treatment step:

   The purpose of PAC contacting was for removal of soluble organics only. The granular media filter was considered part of this treatment step since it removed final traces of the PAC after contacting and gravity sedimentation.

3. Spent powdered activated carbon regeneration system:

   A major objective of this study was to determine the effectiveness of powdered carbon regeneration and reuse. Thus, the preliminary results obtained are presented and discussed separately.

4. Typical overall treatment:

   To allow comparison of PAC-PCT effectiveness with other treatment approaches and provide an indication of effluent quality variations, these results are presented separately.

5. Plant operating and equipment problems:

   Problems encountered, which were considered to have had an effect on treatment results or recommended prototype design, are presented and discussed.

Pilot plant operating and treatment results will be presented as average values with the exception of the following data which will be presented as median values (that value for which fifty percent of the observed values were less than or equal to):

1. Total and soluble TOC, COD and P
2. Suspended solids
3. Solids-contact units reaction zone and blowdown solids concentration
4. Solids-contact units sludge blowdown volumes (percent of flow)
5. Alum and polymer dosages

The chemical and powdered carbon operating and treatment results will be presented in tabular form for different test runs. The runs will be broken down into chemical type and dosages, carbon dosage and number of contacting stages. Since more than 6000 pieces of data were compiled, only selected data are presented in total. These selected data will be SS and P for some chemical treatment runs, SS and SCOD for some carbon treatment runs and SS, TOC, COD and P for typical plant effluent results.

CHEMICAL TREATMENT STEP

Ferric Chloride Treatment Results

After pilot plant start-up and shake down, evaluation of ferric chloride in the chemical treatment step began. About 3 1/2 months of operation followed. The operating and process treatment results for the four Runs conducted are presented in Table 2.

Run F-1 was the first plant test conducted and minimal data was compiled, compared to other Runs. Run F-2 was conducted using operating conditions similar to Run F-1 except that a different polyelectrolyte and a high $FeCl_3$ dosage was used. During Run F-3 higher dosages of $FeCl_3$ and polyelectrolyte were used at a reduced overflow rate for the purpose of establishing the best quality effluent obtainable by chemical treatment with $FeCl_3$. Excellent COD, SS and total phosphorus removals were achieved. However, chemical costs alone ($FeCl_3$ and polyelectrolyte) amounted to about 7¢/1000 gallons. Run F-4, only four days long, was conducted at a reduced overflow rate to obtain an indication of whether or not

TABLE 2: $FeCl_3$ TREATMENT: SUMMARY OF RESULTS

OPERATING RESULTS:

Run Number	F-1	F-2	F-3	F-4
Length of run, days	52	37	13	4
Overflow rate, gpm/sq ft	0.75	0.75	0.50	0.25
$FeCl_3$, mg/ℓ	99	119	145	116
Polymer, type	Dow A23[a]	Atlasep 2A2[b]	Atlasep 2A2	--
Polymer, mg/ℓ	1.0	1.0	1.5	0
Reaction zone solids, g /ℓ	--	2.9	3.3	1.2
Blowdown Volume, % of flow	--	0.9	1.3	2.0
Blowdown Conc., g /ℓ	--	17	13	10
Blowdown solids, lb /MG	--	1280	1410	1670

PROCESS TREATMENT RESULTS:

	INF.	EFF.	INF.	EFF.	INF.	EFF.	INF.	EFF.
Turbidity, JTU[c]	36	24/9	33	16/10	52	26/5	59	42/5
Suspended solids, mg/ℓ	96	30	75	12	94	14	138	33
Total phosphorus, mg/ℓ	5.0	0.7	6.0	0.9	7.8	0.2	7.3	0.5
TOC, mg/ℓ	62	32	60	30	77	21	64	22
COD, mg/ℓ	--	--	121	38	255	54	271	65
pH, units	--	--	--	--	7.6	6.9	7.5	6.9
Total iron, mg/ℓ as Fe	--	--	--	--	1.2	6.2	1.9	8.5
Temperature, °C	--	--	18	--	21	--	21	--

a = A product of Dow Chemical Company, Midland, Michigan
b = A product of Atlas Chemical Company, Willmington, Delaware
c = Effluent Turbidity - 24 Hr. composite/average of 8-10 grabs samples per day

polyelectrolyte flocculation aid could be eliminated without loss of clarification effectiveness. Effective clarification and phosphorus removal was accomplished without the addition of polyelectrolyte.

The variability of feed and effluent suspended solids and phosphorus during Run F-2 is shown in Figure 21. Feed and effluent suspended solids varied by an order of magnitude (i.e., a factor of 10). The effluent phosphorus concentration also varied over an order of magnitude, whereas, the feed concentration was more uniform, varying only by a factor of about 2 from low to high values measured.

The test results for $FeCl_3$ treatment in Table 2 do not show any quantitative correlations between clarification effectiveness and overflow rate or chemical dosage. As would be expected, however, the data does generally indicate that the phosphorus residual is inversely proportional to $FeCl_3$ dosage and directly related to chemical treatment clarifier effluent grab samples turbidity (i.e., clarification effectiveness).

Some 10-15 percent of the iron fed to the raw wastewater appeared in the chemical treatment clarifier effluent during Runs F-3 and F-4. Figure 22 shows that at higher overflow rates, or, conversely, shorter liquid detention times, the ratio of effluent composite sample turbidity to instantaneous, or grab-sample, turbidity decreased substantially. Either the nature of the suspended solids, as measured by turbidity, changed during sample storage or post precipitation of some compound was occurring. The solids present in stored samples had a reddish-brown color and resembled ferric floc.

It has been indicated by others that the iron phosphorus material, precipitated by addition of ferric iron to an aerobic phosphorus bearing wastewater, will release phosphorus when subjected to an anaerobic environment[24,25]. These investigators conjectured that the ferric iron was reduced to ferrous iron in the anaerobic environment.

FIGURE 21: $FeCl_3$ TREATMENT: FEED AND EFFLUENT SUSPENDED SOLIDS AND PHOSPHORUS

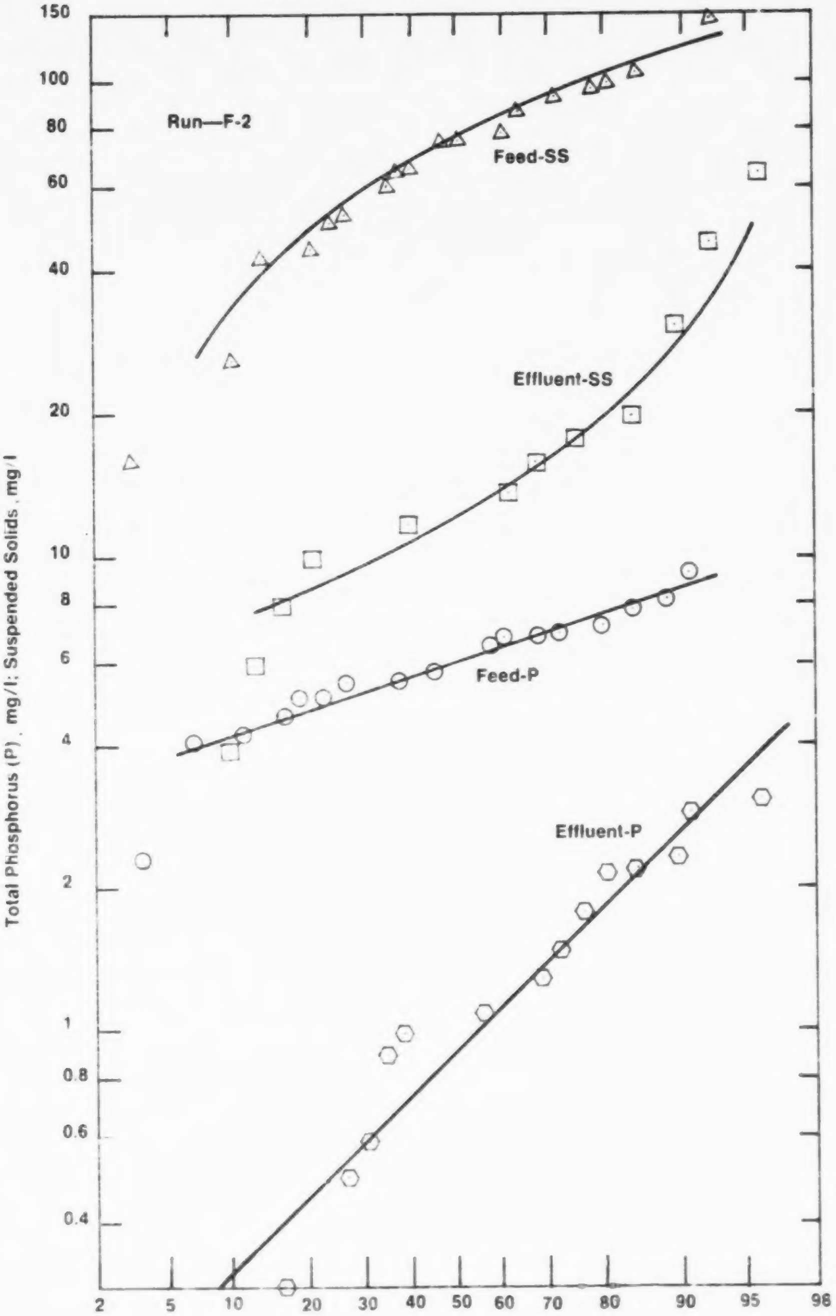
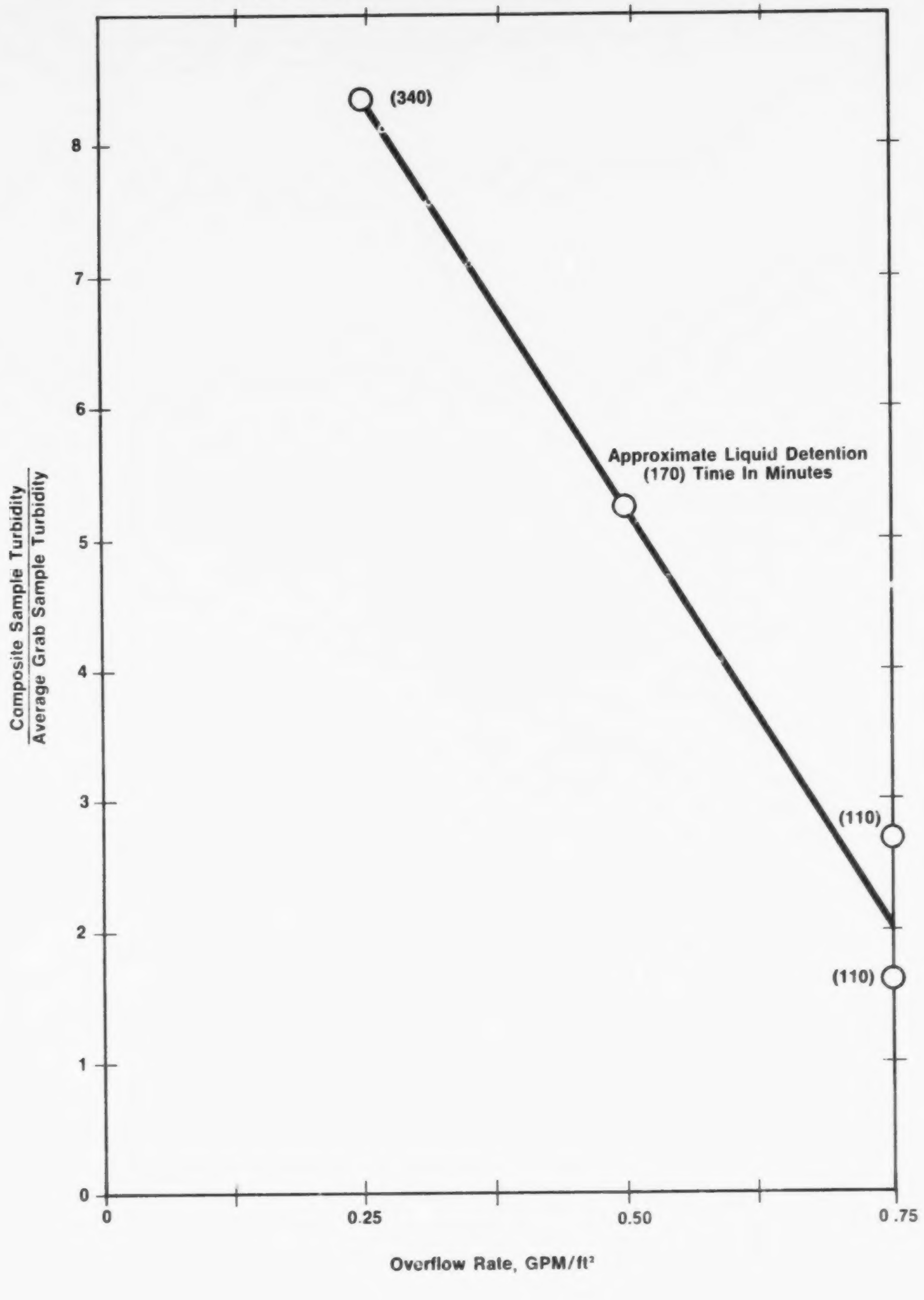


FIGURE 22: FeCl$_3$ TREATMENT: POST-PRECIPITATION OF IRON



It is probable that addition of ferric iron to the raw wastewater, which was devoid of dissolved oxygen, resulted in some ferric iron being reduced to ferrous iron. If, at some later time, this wastewater adsorbed oxygen, the ferrous iron could have been oxidized to ferric iron and precipitated as hydrated ferric oxide (iron flow).

The data in Figure 22 show that long liquid detention times result in high levels of turbidity increases in stored (composite) samples. Operation of the solids-contact unit entailed maintenance of an approximately constant inventory of sludge within the unit to accomplish solids-contacting and some sludge thickening. Thus, the longer the liquid detention time (i.e., low overflow rate) or the lower the solids production rate, the greater will be the average sludge age. During Run F-4 at about 5 1/2 hour liquid detention time, the sludge blowdown was considerably more odorous and darker in color than previously experienced, indicating septic conditions. It is probable that the greater the anaerobic activity of the sludge the more pronounced is the problem of turbidity increases in stored chemical clarifier effluent samples.

Sludge Production

Sludge production can be determined from the data in Table 2. The blowdown solids represent the median amount of sludge removed from the wastewater via the chemical clarifier underflow, including both raw wastewater solids and chemical solids (i.e., ferric floc).

Table 3 is a summary of sludge production data for ferric chloride treatment Runs F-2, 3 and 4. It is seen that between 0.6 and 0.8 pounds of chemical solids (ferric floc) was produced for each pound of $FeCl_3$ fed to the wastewater. A value of 0.87 pounds chemical sludge produced per pound $FeCl_3$ used has been reported elsewhere[26].

Sludge Thickening and Dewatering

The chemical treatment unit underflow concentration ranged from 10 to 17 g/ℓ, or about 1.0 to 1.7 percent dry solids by weight (Table 2). This concentration range was somewhat

TABLE 3: $FeCl_3$ TREATMENT: SUMMARY OF SLUDGE PRODUCTION DATA

Run #	F-2	F-3	F-4
Blowdown SS, lb/MG[a]	1280	1410	1670
Blowdown SS, mg/ℓ[a]	154	169	200
Clarifier Effluent SS, mg/ℓ	12	14	33
Total Chemical and Sewage SS, mg/ℓ	166	183	233
Raw Wastewater SS, mg/ℓ	75	94	138
Chemical SS, mg/ℓ	91	89	95
$FeCl_3$ Dosage, mg/ℓ	119	145	116
lb Chemical Solids/lb $FeCl_3$ Fed	0.77	0.61	0.82
% Chemical Solids in Blowdown Sludge[b]	55	49	40

a = Based on total plant flow
b = Assuming clarifier effluent suspended solids are one-half chemical and one-half raw wastewater solids.

less than anticipated for the approximately 1:1 ferric-primary sludges. The total-solids circulation feature of the solids-contact treatment unit employed may have precluded effective sludge thickening within the chemical treatment unit.

No attempt was made to maximize underflow concentration by varying sludge blowdown duration and/or frequency. The actual duration and frequency of blowdown used was determined by the plant operators on the basis of controlling the sludge pool depth at about two feet above the sludge rakes which insured solids contacting. At low hydraulic loadings, less concentrated blowdown solids were produced and therefore, larger blowdown volumes were required to remove the sludge produced. Recall that low hydraulic loadings are correlated with long sludge age for the solids-contact units used. It is possible that the older the sludge (longer sludge age), the poorer it would thicken in the solids-contact unit.

Laboratory sludge thickening tests were conducted on three 24-hour composite samples of chemical-sewage sludge collected during Runs F-3 and 4. Results of these tests indicated that this sludge would gravity thicken to a maximum concentration of 3 to 4 percent solids by weight. Again, this concentration was less than was anticipated for the 1:1 ferric-primary sludge mixture experienced. Figure 23 presents thickener solids loading and underflow solids concentration predicted from laboratory test results for different sludge samples. Based on judgment and experience, an operating line was drawn which would result in a high degree of sludge thickening. A plot of thickener loading and underflow concentration at the intersection of each curve in Figure 23 with the operating line is plotted in Figure 24. These plots represent suggested design data.

A weighted average of chemical clarifier underflow data in Table 2 indicates that about 1.5 percent solids were produced. A conservative extrapolation of the data in Figure 24 would result in prediction of a thickener loading and underflow concentration of about 12 lb/day-sq ft and 4.5 percent solids.

Vacuum filtration leaf tests were conducted on three partially thickened 24-hour composite sludge samples of chemical clarifier underflow. The results of these tests, summaried in Table 4, indicate that sludges of from 1.6 to 3.3 percent concentration were dewatered

FIGURE 23: $FeCl_3$ TREATMENT:
LABORATORY SLUDGE THICKENING RESULTS

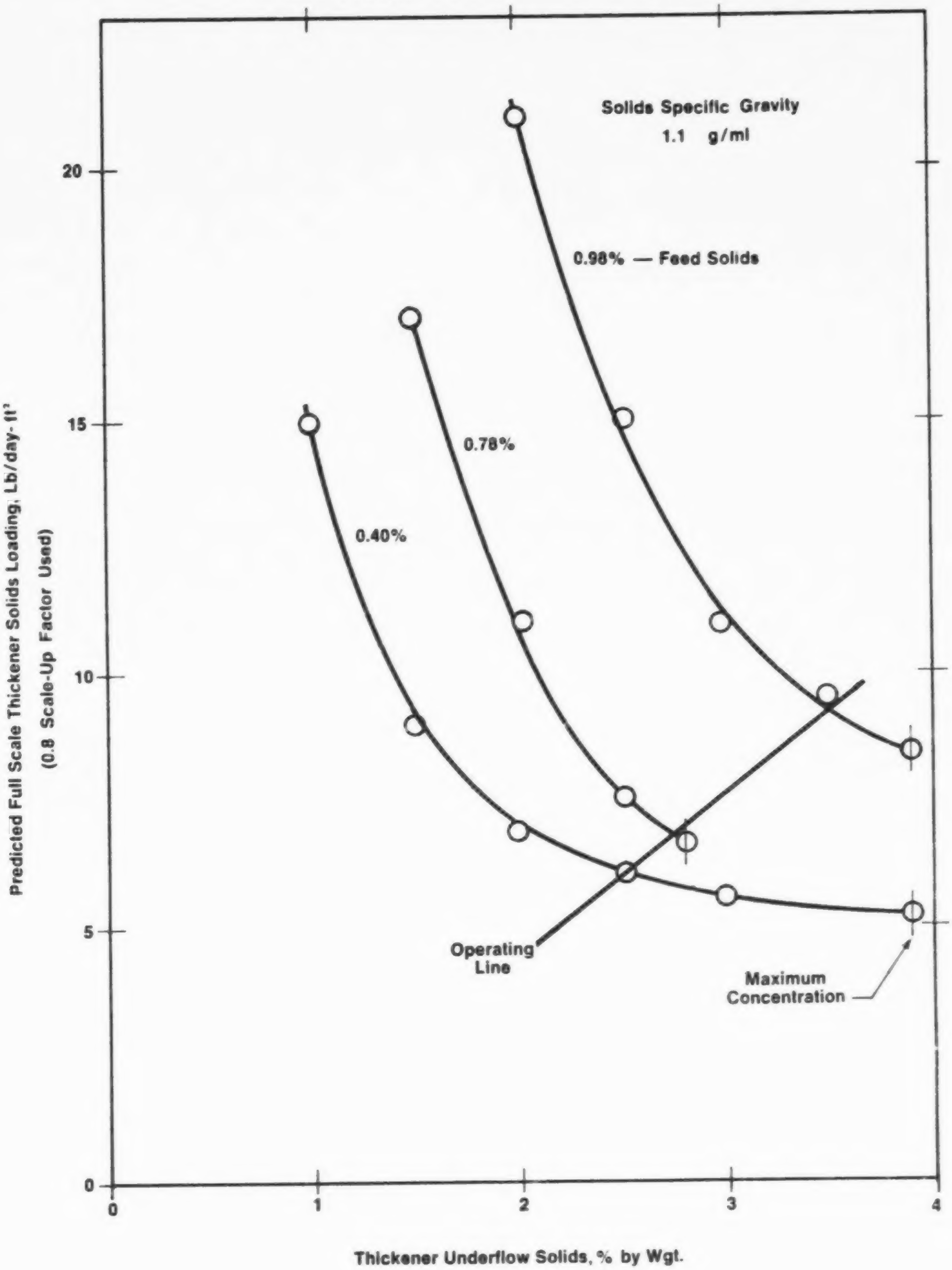


FIGURE 24: $FeCl_3$ TREATMENT: SLUDGE THICKENING

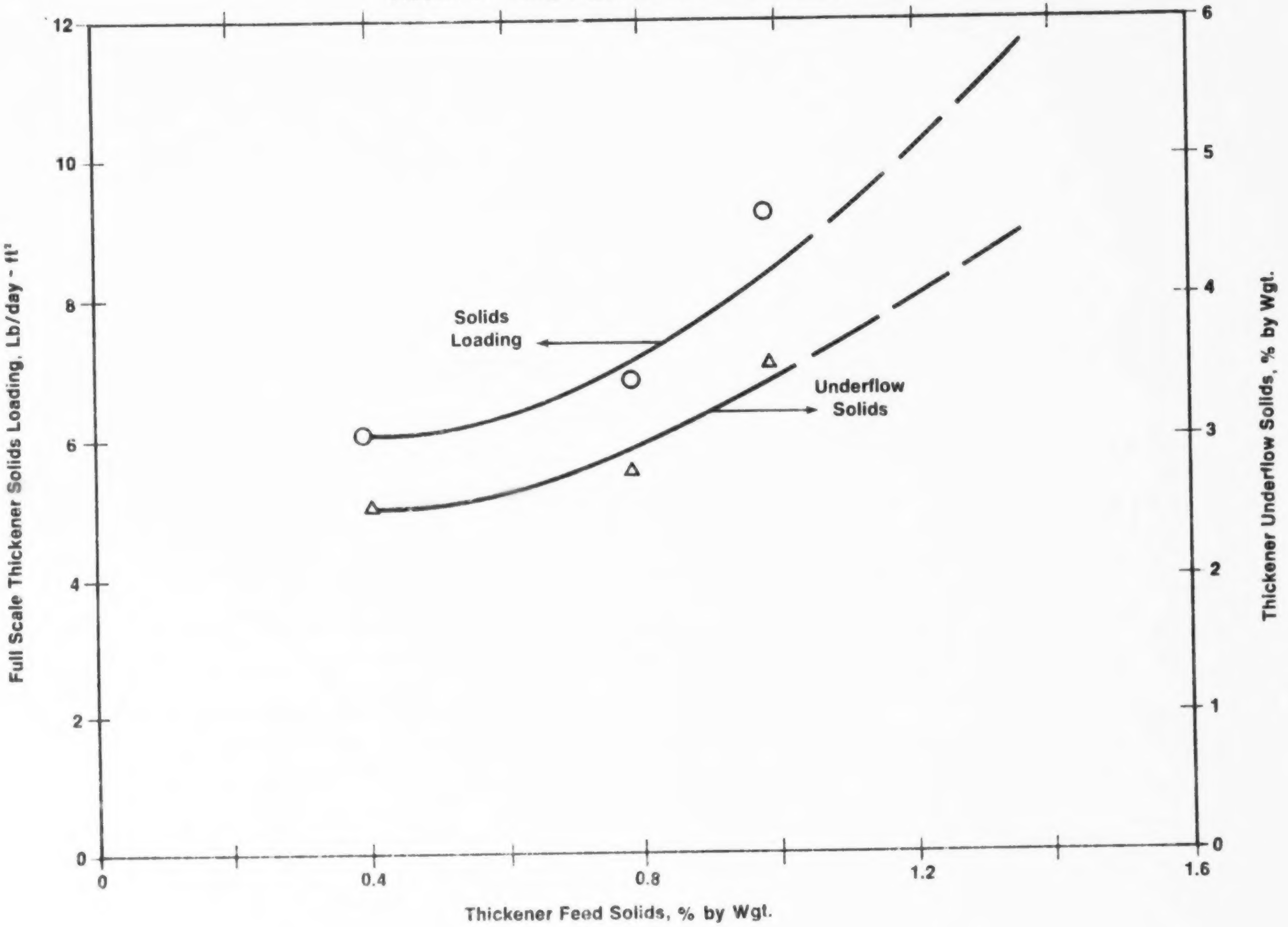


TABLE 4: $FeCl_3$ TREATMENT: VACUUM FILTRATION LEAF TEST RESULTS

Test	Feed Solids % by Weight	Cycle Time Minutes	Full Scale Filtration Yield[a] lb/hr-sq ft	Cake Moisture % by Weight
1	2.5	16	0.37	85
2	3.3	20	0.32	83
3	1.6	25	0.27	80
4,5	1.6[b]	5	1.0	82
6	1.6[c]	5.3	1.0	83

a = Maximum possible yield of dry solids (based on feed and chemical solids) filter cake, at 33% submergence and a 0.8 scale-up factor.

b = 24% $Ca(OH)_2$ by weight added for conditioning.

c = 47% $Ca(OH)_2$ by weight added for conditioning.

to 80-85 percent moisture at predicted full scale filter yields of about 0.3 pounds of dry solids per hour per square foot of filter area. These filter yields were predicted for a 3/16 inch filter cake thickness which was only marginally dischargable. Conditioning of the dilute ferric-sewage sludge with 24 percent lime on a dry solids basis resulted in a dischargable 3/16 inch filter cake and a filter yield of 1 lb/hr-sq ft.

The thickening and vacuum dewatering test results reported are considered to be conservative. Additional studies of ferric-sewage sludge handling and chemical conditioning are necessary to achieve reliable design data.

Alum Treatment Results

Two brief comprehensive Runs were conducted using liquid alum to effect coagulation of suspended solids and phosphorus precipitation. A summary of operating and process treatment results is presented in Table 5. Runs A-1 and A-2 were conducted using identical alum dosages. During Run A-1 no polyelectrolyte was added while during Run A-2, 0.25 mg/ℓ of anionic polyelectrolyte was added. The purpose of these comparative Runs was to determine the increased hydraulic loading possible when using polyelectrolyte flocculation aid. The dosage and type of polyelectrolyte were chosen on the basis of jar test results. The type used was the most effective of four anionic polyelectrolytes evaluated and the dosage near optimum based on floc settling rates. It was assumed that a raw wastewater phosphorus concentration of 6-7 mg/ℓ would be experienced and could be reduced to less than 1.0 mg/ℓ at an aluminum to phosphorus molar ratio of 2:1. Therefore, an alum dosage of about 140 mg/ℓ was used.

The operating results in Table 5 show that use of 0.25 mg/ℓ polyelectrolyte allowed an increase in overflow rate from 0.35 gpm/sq ft without polyelectrolyte, to 0.60 gpm/sq ft. Both of these overflow rates were about 80 percent of the maximum rates possible to maintain reasonably stable operating and effluent quality conditions.

Process treatment results in Table 5 show that near identical effluent quality was produced during the two runs. A similar variation of raw wastewater and chemical treatment unit effluent suspended solids and phosphorus values is shown in Figure 25.

TABLE 5: ALUM TREATMENT: SUMMARY OF RESULTS

OPERATING RESULTS:

Run #	A-1	A-2
Length of Run, days	13	12
Overflow rate, gpm/sq ft	0.35	0.60
Al^{+3}/Filter Alum[a], mg/ℓ	13/143	13/143
Polymer, type	--	Atlasep 3A3[b]
Polymer, mg/ℓ	0	0.25
Reaction zone solids, g /ℓ	0.8	0.6
Blowdown volume, % of flow	6.2	3.5
Blowdown conc., g /ℓ	1.8	4.4
Blowdown solids, lb/MG	930	1280

PROCESS TREATMENT RESULTS:

	INF.	EFF.	INF.	Eff.
Turbidity, JTU	36	10	38	11
Suspended solids, mg/ℓ	82	24	110	30
Total phosphorus, mg/ℓ	4.4	1.0	5.8	1.2
Soluble phosphorus, mg/ℓ	1.5	0.4	3.0	0.4
TOC, mg/ℓ	73	28	70	23
STOC, mg/ℓ	24	19	26	15
COD, mg/ℓ	155	66	158	45
SCOD, mg/ℓ	40	30	67	27
BOD_5, mg/ℓ	133	30	94	38
pH, units	7.3	7.1	7.4	7.0
Temperature, °C	17	--	16	--

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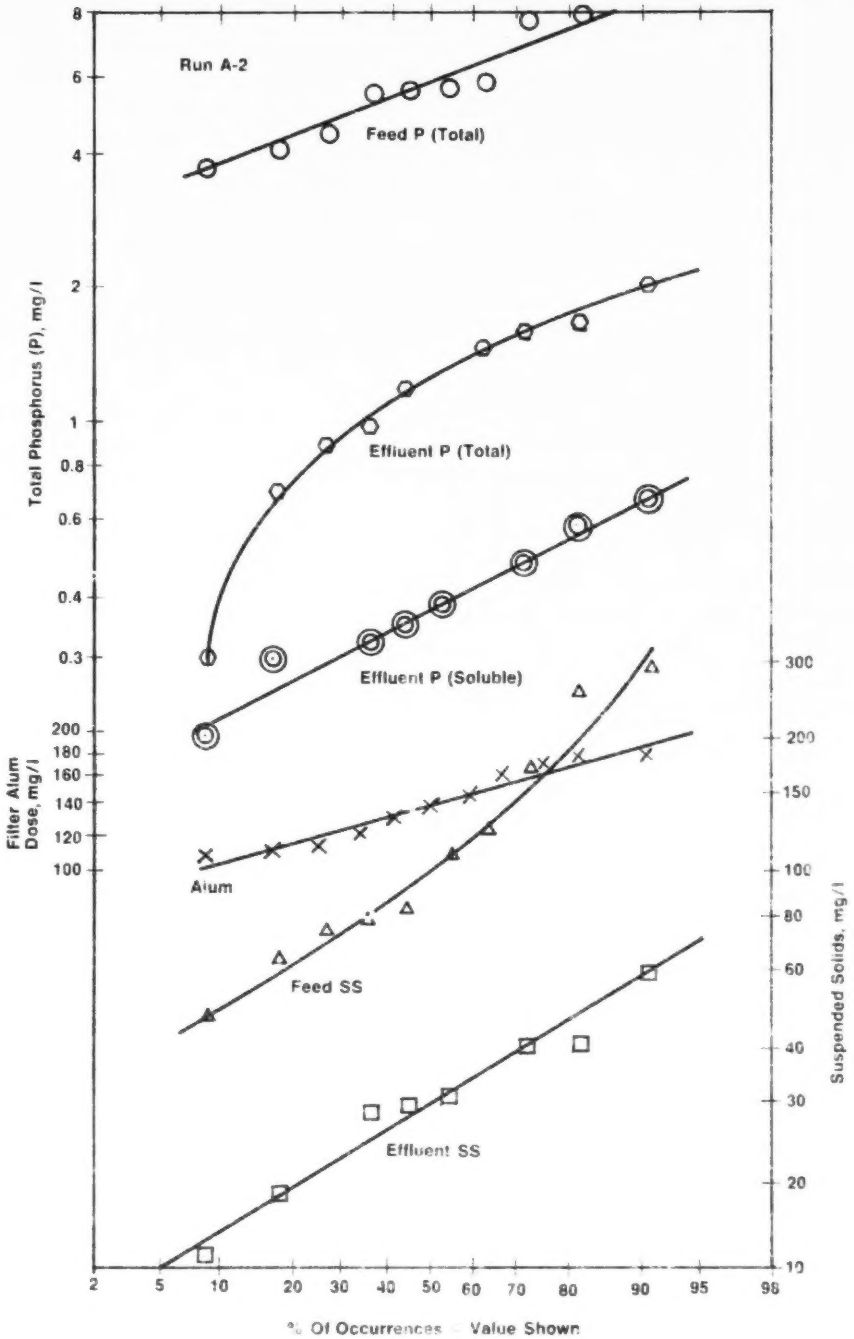
a - $Al_2(SO_4)_3 \cdot 14\ H_2O$ = Filter Alum
b - A product of Atlas Chemical Company, Wilmington, Delaware

FIGURE 25: ALUM TREATMENT: DOSAGE, FEED AND EFFLUENT SUSPENDED SOLIDS AND PHOSPHORUS



One aspect of the cost effectiveness of using polyelectrolyte can be determined by comparing the operating cost for using polyelectrolyte versus the capital cost of buying increased clarification area. Table 6 shows such a comparison for alum treatment. The estimated installed cost of solids-contact treatment equipment to treat 10 mgd flow without polyelectrolyte is 0.47¢/1000 gallon. If polyelectrolyte is used, the tabulated values show the overflow rate required to justify use of polyelectrolyte on the basis of holding total treatment cost constant. It is indicated that to justify use of 0.15 mg/ℓ of polyelectrolyte, the overflow rate would have to be increased more than five fold (i.e., from 0.27 to 1.46 gpm/sq ft). For alum treatment it would appear more economical to design on the basis of operating without polyelectrolyte. However, since use of polyelectrolyte does substantially increase floc settling rate, it would be recommended that polymer make-up and feed facilities be provided to increase treatment flexibility and reliability.

Based on effluent turbidity values of 10 and 11 JTU shown in Table 5, good clarification was experienced. No difference was observed between the turbidities of 24-hour composite samples and the average turbidity of 8 to 10 grab samples per day. The effluent suspended solids values of 24 and 30 mg/ℓ were higher than the desired 10 to 20 mg/ℓ. Presumably effluent turbidities of 4 to 6 JTU should have been strived for rather than the 1-15 JTU level set as an operating criteria.

Comparison of effluent total and soluble phosphorus values indicates that 60 to 67 percent of the chemical clarifier effluent total phosphorus was particulate (i.e., the difference between total phosphorus and soluble phosphorus). After subsequent treatment in the powdered carbon contactors and the granular media filter, the total phosphorus in the plant effluent was identical to the soluble phosphorus in the chemical clarifier effluent. Thus, complete removal of particulate phosphorus was obtained for the PCT plant. The low phosphorus value of 0.4 mg/ℓ was achieved at aluminum to total phosphorus molar ratios of 2.6:1 and 3.4:1.

Of significance was the removal of 20 and 24 percent soluble TOC and 25 and 60 percent SCOD across the chemical treatment step. Obviously, alum treatment coagulated some organic material (soluble or colloidal) which would pass a 0.45 micron membrane filter. This observation verifies the laboratory jar test data using alum where some 10 to 20 percent removal of SCOD was observed.

TABLE 6: ALUM TREATMENT: COST TRADE-OFF OF POLYELECTROLYTE AND CLARIFICATION AREA FOR 10 MGD FLOW

SOLIDS-CONTACT UNIT		POLYELECTROLYTE		
Overflow Rate (gpm/sq ft)	Amortized Capital Cost[a] (¢/1000 gal.)	Dose (mg/ℓ)	Chemical Cost[b] (¢/1000 gal.)	Total Cost (¢/1000 gal.)
0.27	0.47	0	0	0.47
0.48	0.39	0.06	0.08	0.47
0.97	0.33	0.11	0.14	0.47
1.46	0.28	0.15	0.19	0.47

a - 7 1/2% interest, 20 year for estimated erected equipment

b - @$1.50/lb of polymer

It should be noted that during the alum treatment Runs, the raw wastewater strength was weaker than usual. This condition does not negate the effectiveness of alum treatment for clarification of raw wastewater. Overall chemical treatment process results for Runs A-1 and A-2 indicate removals of 82 percent particulate TOC, 75 percent particulate COD and 72 percent suspended solids.

Sludge Production

A summary of alum treatment sludge production data is presented in Table 7. Between 0.4 and 0.5 pounds of chemical sludge was produced for each pound of alum fed to the raw wastewater. This range is somewhat higher than 0.25 to 0.40 lb/lb reported by others[26].

Sludge Thickening and Dewatering

Alum-sewage sludge thickening and dewatering properties were not at all impressive. To begin with, a concentrated reaction-zone suspension was never established in the solids-contact unit. Solids concentrations of 0.8 and 0.6 g/ℓ were not considered to constitute solids-contact treatment. A concentration of about 2 g/ℓ would have been considered acceptable.

The alum-sewage sludge was not effectively thickened within the chemical contactor. During Run A-1 the blowdown solids were just about double the concentration of the reaction-zone solids concentration. During Run A-2, when polyelectrolyte was used, a seven fold increase in blowdown solids concentration was achieved. Evidently, the total-solids circulation feature of the solids-contact unit used was ineffective for alum-sewage sludge thickening. Because of very high blowdown volumes employed (3.5 to 6.2 percent of total flow), solids were withdrawn not only from the sludge thickening cone but also from the reaction-zone. Use of shorter blowdown durations would have eliminated solids removal from the reaction-zone, resulting in a more concentrated underflow. From the blowdown volume and concentration data in Table 5, it is apparent that use of polyelectrolyte results in a more concentrated underflow. The economic impact of more concentrated underflow on sludge thickening and dewatering was not evaluated.

TABLE 7: ALUM TREATMENT: SUMMARY OF SLUDGE PRODUCTION DATA

Run #	A-1	A-2
Blowdown SS, lb/MG[a]	930	1280
Blowdown SS, mg/ℓ[a]	112	154
Clarifier Effluent SS, mg/ℓ	24	30
Total Chemical and Sewage SS, mg/ℓ	136	184
Raw Wastewater SS, mg/ℓ	82	110
Chemical SS, mg/ℓ	54	74
lb Chemical SS/lb Alum Fed	0.38	0.52
% Chemical Solids in Blowdown Sludge[b]	40	40

---

a - Based on total plant flow

b - Assuming clarifier effluent suspended solids are one-half chemical and one-half raw wastewater solids

During Run A-1, two laboratory thickening tests were conducted on 24-hour composite samples of alum-sewage sludge. The test results are shown in Figure 26. The maximum solids concentration obtained was 2 to 3 percent dry solids by weight. To produce a thickened sludge of 1.5 percent from a 0.36 percent feed, a predicted thickener solids loading of less than 6 lb/day-sq ft is indicated. Of significance is that an increase in thickener feed solids of 40 percent (from 0.26 to 0.36 percent) would allow a three fold increase in thickener solids loading. Though none were evaluated, the use of a polymer sludge conditioning chemical should improve the alum-sewage thickening properties.

Vacuum filtration leaf test data indicated that dilute alum-sewage sludges, of from 0.5 to 1.0 percent dry solids by weight, would not effectively dewater or produce a dischargable cake without chemical conditioning.

Leaf test data presented in Table 8 indicates that the alum-sewage sludge can be dewatered to 77 to 85 percent moisture, albeit at very low filter yields and high conditioning chemical costs (approximately 1/3 to 1/2¢/1000 gallon of raw wastewater). The filter yields shown in Table 8 can reasonably be increased proportionally to anticipated thickener solids underflow. For example, when using an anionic polymer, if a vacuum filtration feed solids of 1.8 percent were to be dewatered, a yield of about 1 lb/hr-sq ft would be predicted. Obviously, more alum-sewage sludge thickening and dewatering data are required to reliably predict full-scale performance.

Lime Treatment Results

Eleven runs, ranging from 8 to 25 days duration, were conducted using lime treatment for phosphorus precipitation and coagulation of raw wastewater suspended solids. A summary of operating and treatment results is presented in Table 9. The reason for conducting more runs with lime treatment than for alum or ferric chloride was that, with lime, considerably higher hydraulic loadings were possible and more stable operation of the solids-contact treatment unit was experienced. When operating the powdered carbon contactors in parallel rather than series, twice as much chemically clarified effluent was required, thus, lime pretreatment was used.

FIGURE 26: ALUM TREATMENT: LABORATORY SLUDGE THICKENING RESULTS

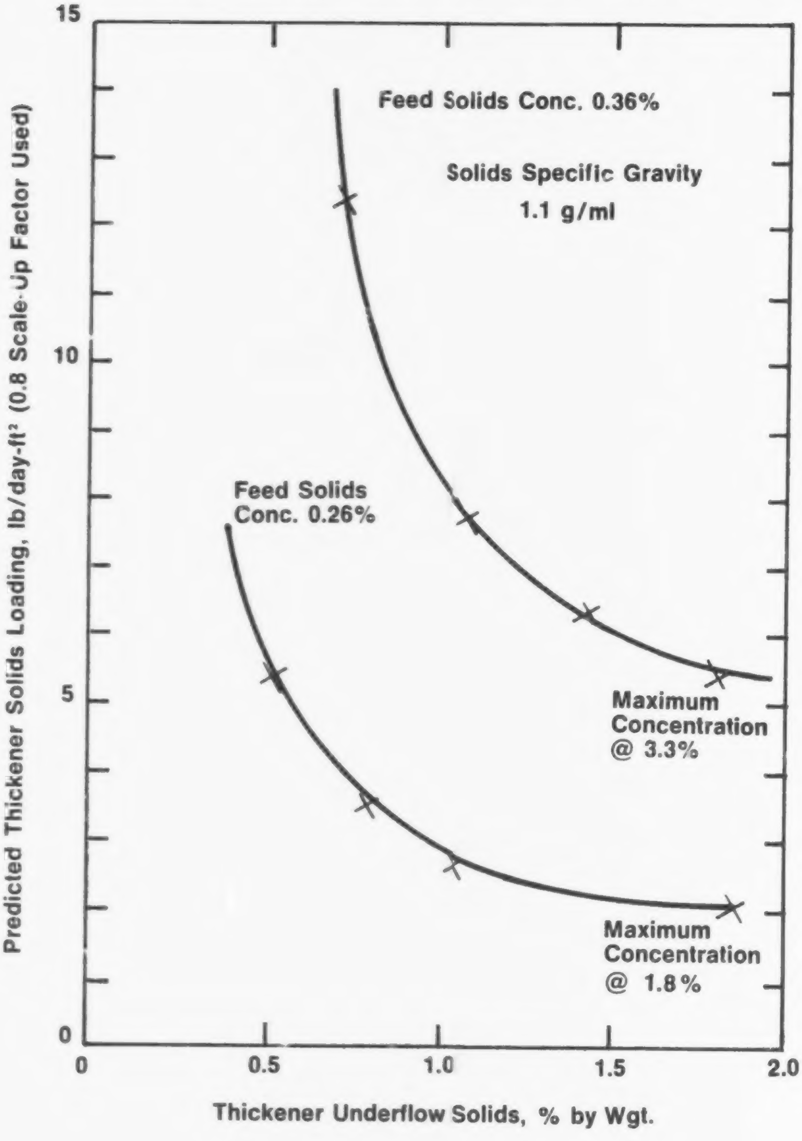


TABLE 8: ALUM TREATMENT: VACUUM FILTRATION LEAF TEST DATA

Test #	1	2
Feed Solids, % by wgt	0.70	0.36
Cycle Time, Minutes	8	23
Yield[a], lb /hr-sq ft	0.38	0.14
Cake Moisture, % by wgt	85	77
Polymer Type	anionic	cationic
Polymer Dosage, lb /ton of Dry Solids	8.6	14

---

a - Maximum possible rate for producing a dischargeable 3/16 inch thick cake, @33% submergence and 0.8 scale - up factor

TABLE 9: LIME TREATMENT: SUMMARY OF RESULTS

OPERATING CONDITIONS:

Run #	L-1	L-2	L-3
Length of Run (days)	9	10	16
Overflow Rate, gpm/sq ft	0.2-0.5	0.51	0.50
Lime Dosage, mg/ℓ [$Ca(OH)_2$]	270	470	425
Reaction Zone pH, units	9.8	10.6	10.8
H_2SO_4 Dosage, lb/MG	2000	1800	--
Reaction Zone Solids, g/ℓ	--	--	8.6
Blowdown Concentration, g/ℓ	--	--	190
Blowdown Volume, % of flow	0.73	0.71	0.49
Blowdown Solids, lb/MG[a]	--	--	9170

PROCESS TREATMENT RESULTS

	INF.	EFF.	INF.	EFF.	INF.	EFF.
Turbidity, JTU	46	15	65	5	46	10
Suspended Solids, mg/ℓ	78	17	140	10	100	16
Total Phosphorus, mg/ℓ	4.3	2.4	7.2	1.5	7.2	1.1
Soluble Total Phosphorus, mg/ℓ	3.2	1.7	4.2	0.98	2.6	0.50
TOC, mg/ℓ	60	49	--	--	66	30
STOC, mg/ℓ	25	32	--	--	22	18
COD, mg/ℓ	146	128	190	42	200	62
SCOD, mg/ℓ	55	73	33	33	--	39
BOD_5, mg/ℓ[b]	--	--	--	--	144	47
pH, Units	7.6	6.1	7.3	8.3	7.5	8.0
Temperature, °C	18	--	--	--	23	--
Alkalinity, mg/ℓ as $CaCO_3$	270	69	309	119	275	110
Total Hardness, mg/ℓ as $CaCO_3$	365	446	420	369	383	335
Ca^{++} Hardness, mg/ℓ as $CaCO_3$	212	278	251	318	230	265

a - Blowdown solids are the median of observed values and are not based on median values of blowdown concentration and volume

b - BOD_5 values are for less than 25 percent of the observations for other parameters

TABLE 9 (cont.): LIME TREATMENT: SUMMARY OF RESULTS

OPERATING CONDITIONS:

Run #	L-4	L-5	L-6
Length of Run (days)	20	16	16
Overflow Rate, gpm/sq ft	0.96	1.3	0.4-1.0
Lime Dosage, mg/ℓ [$Ca(OH)_2$]	365	420	390
Reaction Zone pH, units	10.8	10.8	10.9
H_2SO_4 Dosage, lb/MG	1970	1710	3000
Reaction Zone Solids, g /ℓ	--	2.1	1.9
Blowdown Concentration, g /ℓ	100	64	113
Blowdown Volume, % of flow	0.88	0.75	0.45
Blowdown Solids, lb/MG	7010	5250	5840

PROCESS TREATMENT RESULTS

	INF.	EFF.	INF.	EFF.	INF.	EFF.
Turbidity, JTU	52	5	26	4	28	7
Suspended Solids, mg/ℓ	170	8.0	92	13	~56	12
Total Phosphorus, mg/ℓ	6.7	1.3	3.8	0.3	--	--
Soluble Total Phosphorus, mg/ℓ	3.8	0.9	2.2	0.2	--	--
TOC, mg/ℓ	--	--	47	28	49	30
STOC, mg/ℓ	--	--	23	25	22	26
COD, mg/ℓ	240	50	86	52	130	88
SCOD, mg/ℓ	37	41	38	45	55	50
BOD_5, mg/ℓ	68	16	--	--	--	--
pH, Units	7.6	7.6	7.6	7.9	7.5	7.8
Temperature, °C	--	--	15	--	18	--
Alkalinity, mg/ℓ as $CaCO_3$	306	62	336	123	312	89
Total Hardness, mg/ℓ as $CaCO_3$	404	327	465	410	433	406
Ca^{++} Hardness, mg/ℓ as $CaCO_3$	251	265	280	263	270	309

TABLE 9 (cont): LIME TREATMENT: SUMMARY OF RESULTS

OPERATING CONDITIONS:

Run #	L-7	L-8	L-9
Length of Run (days)	23	24	8
Overflow Rate, gpm/sq ft	0.7	0.7-1.0	1.1
Lime Dosage, mg/ℓ [$Ca(OH)_2$]	526	490	586
Reaction Zone pH, units	10.9	11.0	11.0
H_2SO_4 Dosage, lb/MG	2550	2500	3690
Reaction Zone Solids, g /ℓ	3.8	12	2.2
Blowdown Concentration, g /ℓ	81	185	54
Blowdown Volume, % of flow	1.4	0.40	1.5
Blowdown Solids, lb/MG	8340	7340	7260

PROCESS TREATMENT RESULTS

	INF.	EFF.	INF.	EFF.	INF.	EFF.
Turbidity, JTU	50	4	24	7	46	5
Suspended Solids, mg/ℓ	120	7.0	~58	17	92	1.7
Total Phosphorus, mg/ℓ	8.0	1.6	5.3	1.0	4.9	0.98
Soluble Total Phosphorus	4.8	0.96	1.6	0.5	3.0	0.7
TOC, mg/ℓ	75	30	--	--	76	25
STOC, mg/ℓ	24	24	--	--	19	20
COD, mg/ℓ	260	54	90	65	155	53
SCOD, mg/ℓ	35	43	40	44	41	44
BOD_5, mg/ℓ	--	--	57	46	154	27
pH, Units	7.5	7.7	7.8	7.3	7.5	7.6
Temperature, °C	15	--	22	--	15	--
Alkalinity, mg/ℓ as $CaCO_3$	341	63	313	54	364	62
Total Hardness, mg/ℓ as $CaCO_3$	444	367	383	320	503	469
Ca^{++} Hardness, mg/ℓ as $CaCO_3$	255	297	229	265	283	409

TABLE 9 (cont.): LIME TREATMENT: SUMMARY OF RESULTS

OPERATING CONDITIONS:

Run #	L-10	L-11
Length of Run (days)	20	25
Overflow Rate, gpm/sq ft	0.5-0.7	0.4
Lime Dosage, mg/ℓ [$Ca(OH)_2$]	800-1000	1530
Reaction Zone pH, units	11.3-11.6	11.6
H_2SO_4 Dosage, lb/MG	4200-6600	14,300
Reaction Zone Solids, g /ℓ	5.8	4.4
Blowdown Concentration, g /ℓ	54	42
Blowdown Volume, % of flow	1.4	2.8
Blowdown Solids, lb/MG	6300	12,100

PROCESS TREATMENT RESULTS

	INF.	EFF.	INF.	EFF.
Turbidity, JTU	24	7	55	4
Suspended Solids, mg/ℓ	72	6.0	170	6.5
Total Phosphorus, mg/ℓ	3.8	0.5	7.8	1.0
Soluble Total Phosphorus	2.5	0.4	4.5	0.5
TOC, mg/ℓ	--	--	--	--
STOC, mg/ℓ	--	--	--	--
COD, mg/ℓ	120	79	220	40
SCOD, mg/ℓ	44	62	40	35
BOD_5, mg/ℓ	65	47	--	--
pH, Units	7.6	6.8	7.5	7.6
Temperature, °C	21	--	15	--
Alkalinity, mg/ℓ as $CaCO_3$	316	67	330	80
Total Hardness, mg/ℓ as $CaCO_3$	392	543	464	1550
Ca^{++} Hardness, mg/ℓ as $CaCO_3$	216	512	280	1500

Hydraulic loading and treatment pH were the primary operational parameters evaluated.

Run L-1 was conducted at a pH of 9.8. Effluent suspended solids and turbidity were 17 mg/ℓ and 15 JTU. During Run L-1, a reduction of overflow rate from 0.5 to 0.2 gpm/sq ft did not significantly reduce effluent suspended solids. An effluent phosphorus concentration of 2.4 mg/ℓ was observed. The low phosphorus removal at pH 9.8 was not unexpected. Laboratory jar test results had indicated that a higher treatment pH was required to reduce phosphorus to lower levels.

Runs L-2 through L-9 were conducted at moderate treatment pH's ranging from 10.6 to 11.1 and overflow rates ranging from about 0.5 to 1.3 gpm/sq ft. Clarification effectiveness ranged from marginal (10 JTU, 17 mg/ℓ SS and 1.6 mg/ℓ P) to excellent (4 JTU, 2 mg/ℓ SS and 0.3 mg/ℓ P). Runs L-10 and L-11 were conducted at relatively high treatment pH's, above 11.3. Excellent clarification was achieved. Typical feed and effluent suspended solids and phosphorus data for moderate and high treatment pH (10.9 and 11.6) Runs are presented in Figures 27 through 30.

An attempt was made to correlate percent suspended solids removal or concentration of effluent suspended solids with treatment pH or overflow rate. No precise correlation was found with either of these treatment parameters. Recognizing that the coagulation effectiveness of lime treatment is generally attributed to the precipitation of $Mg(OH)_2$, effluent suspended solids were compared with non-calcium hardness ($\propto Mg^{++}$) removals. This correlation was also imprecise.

Figure 31 presents the only precise correlation found between operational parameters and clarification effectiveness. It is seen that poorer effluent quality is associated with increasing sludge age within the chemical treatment unit. The relative sludge age shown is directly related to the reaction-zone suspended solids concentration and inversely related to the hydraulic loading and solids production rate. The actual sludge age was estimated to never have exceeded more than one day.

FIGURE 27: LIME TREATMENT: FEED AND EFFLUENT SUSPENDED SOLIDS

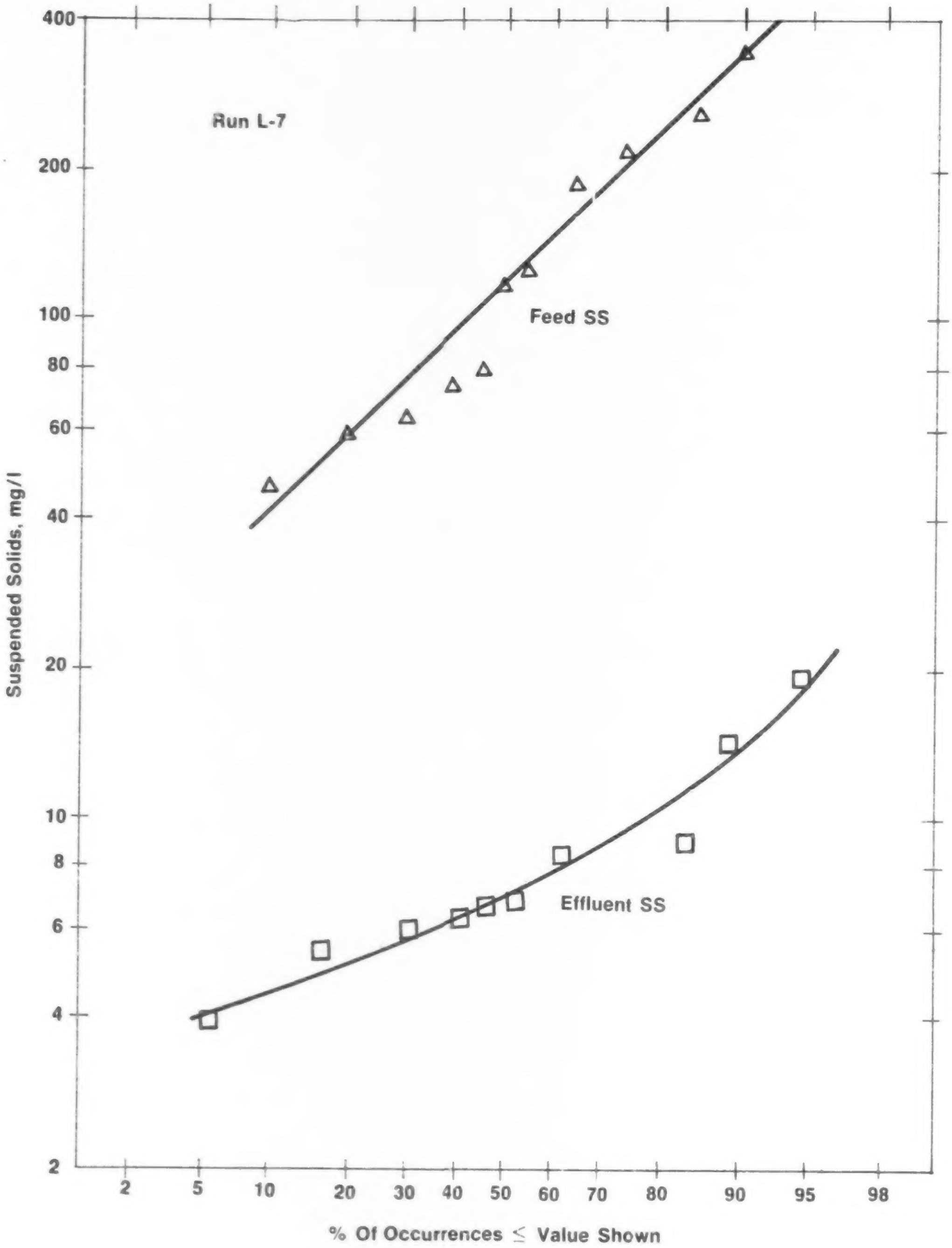


FIGURE 28: LIME TREATMENT: FEED AND EFFLUENT PHOSPHORUS

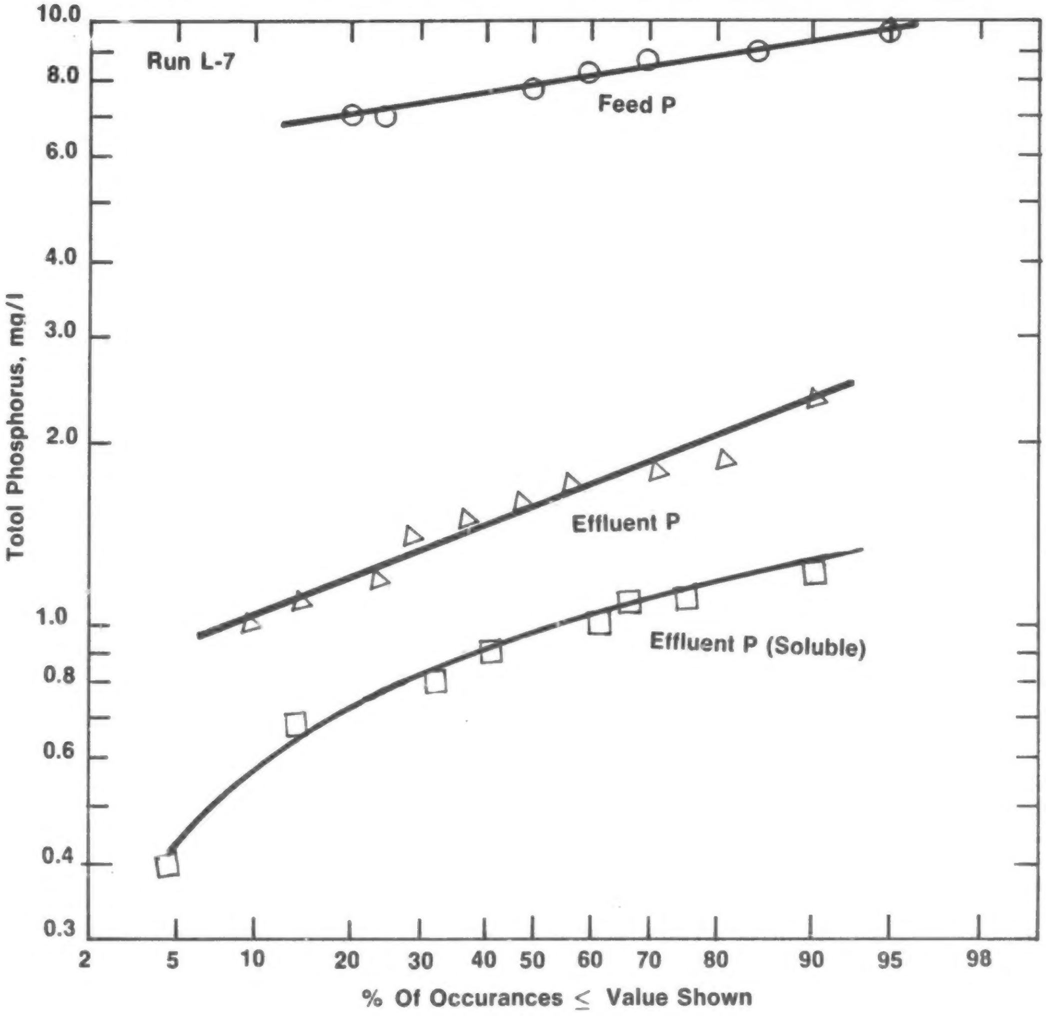


FIGURE 29: LIME TRFATMENT: FEED AND EFFLUENT SUSPENDED SOLIDS

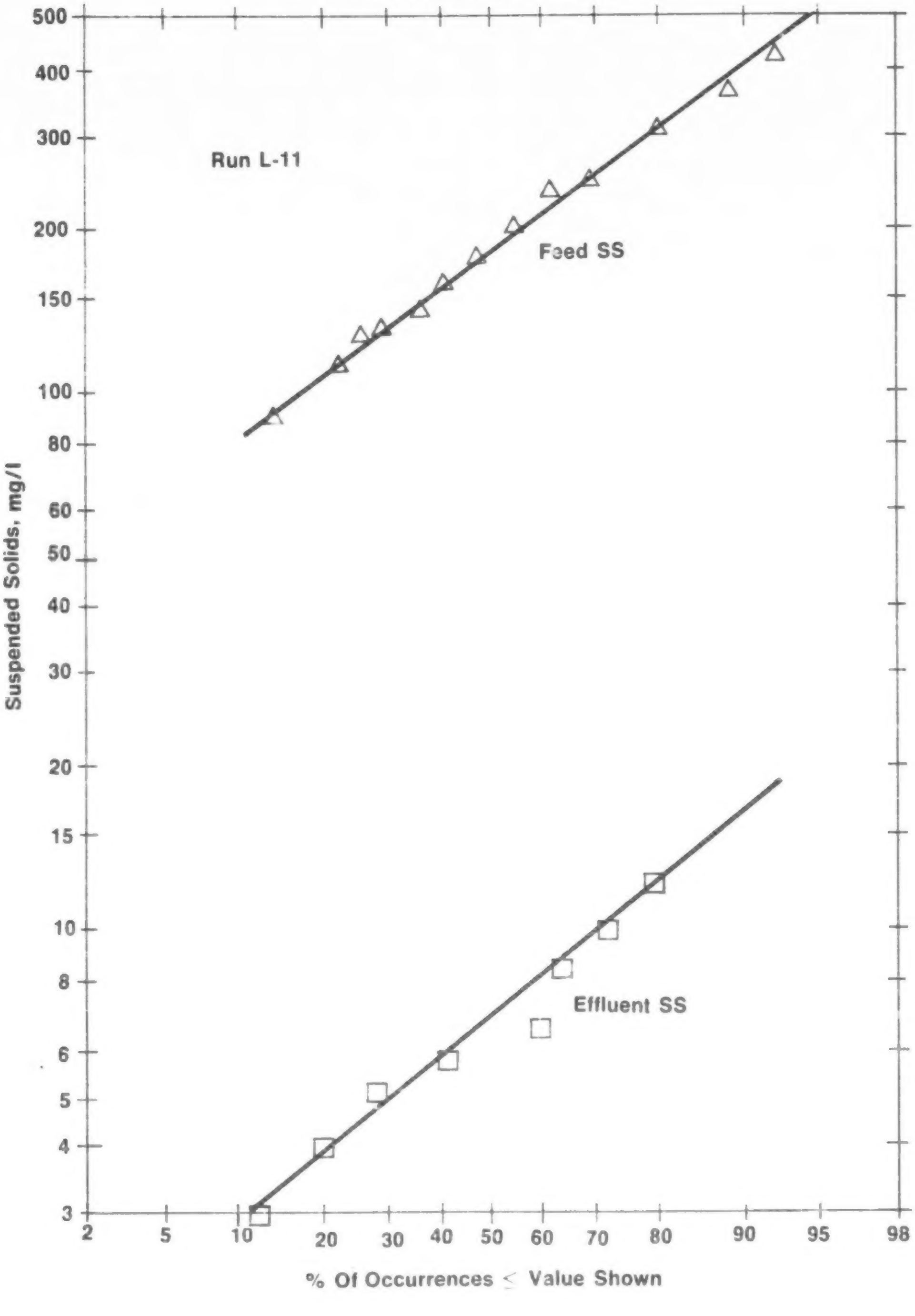


FIGURE 30: LIME TREATMENT: FEED AND EFFLUENT PHOSPHORUS

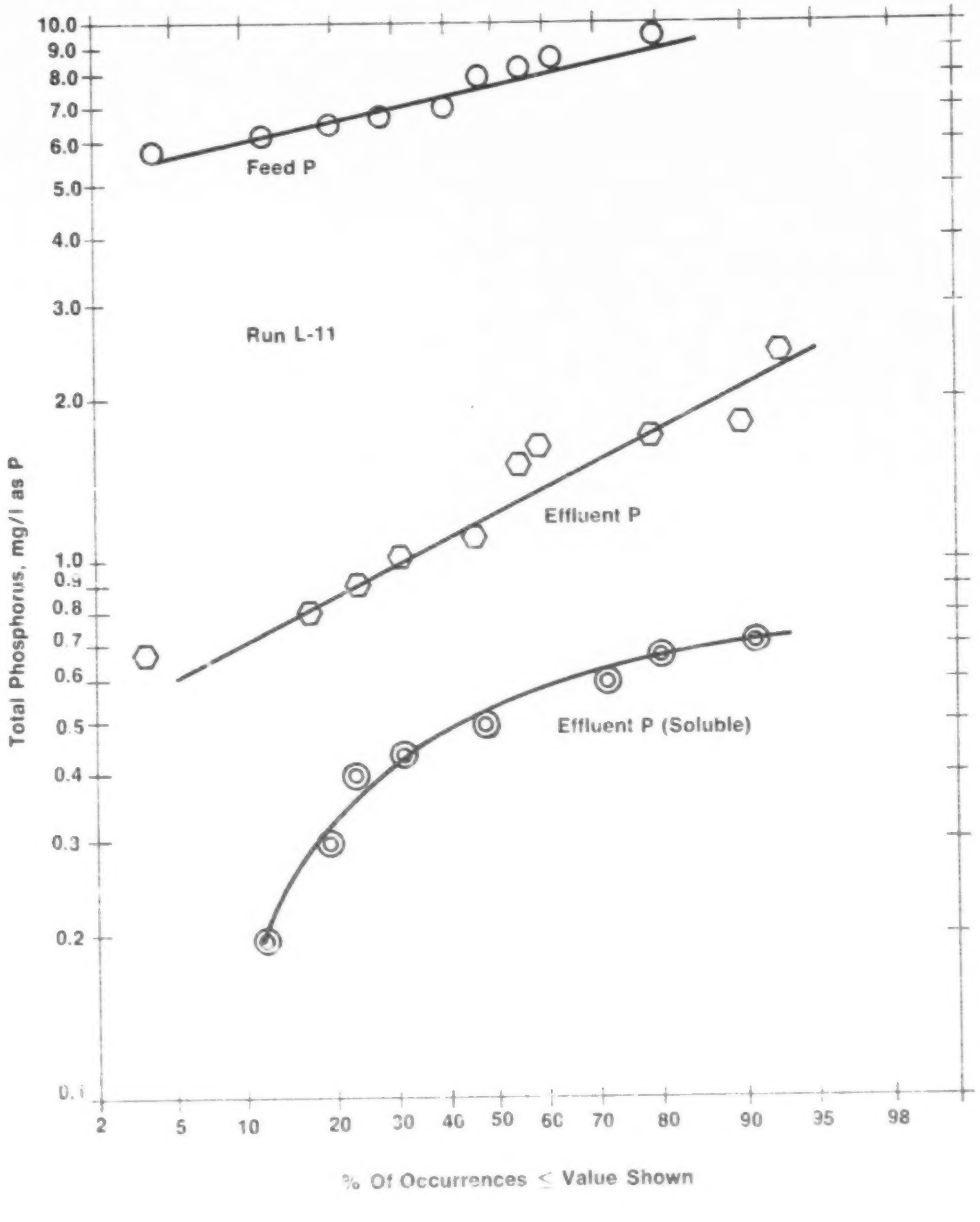
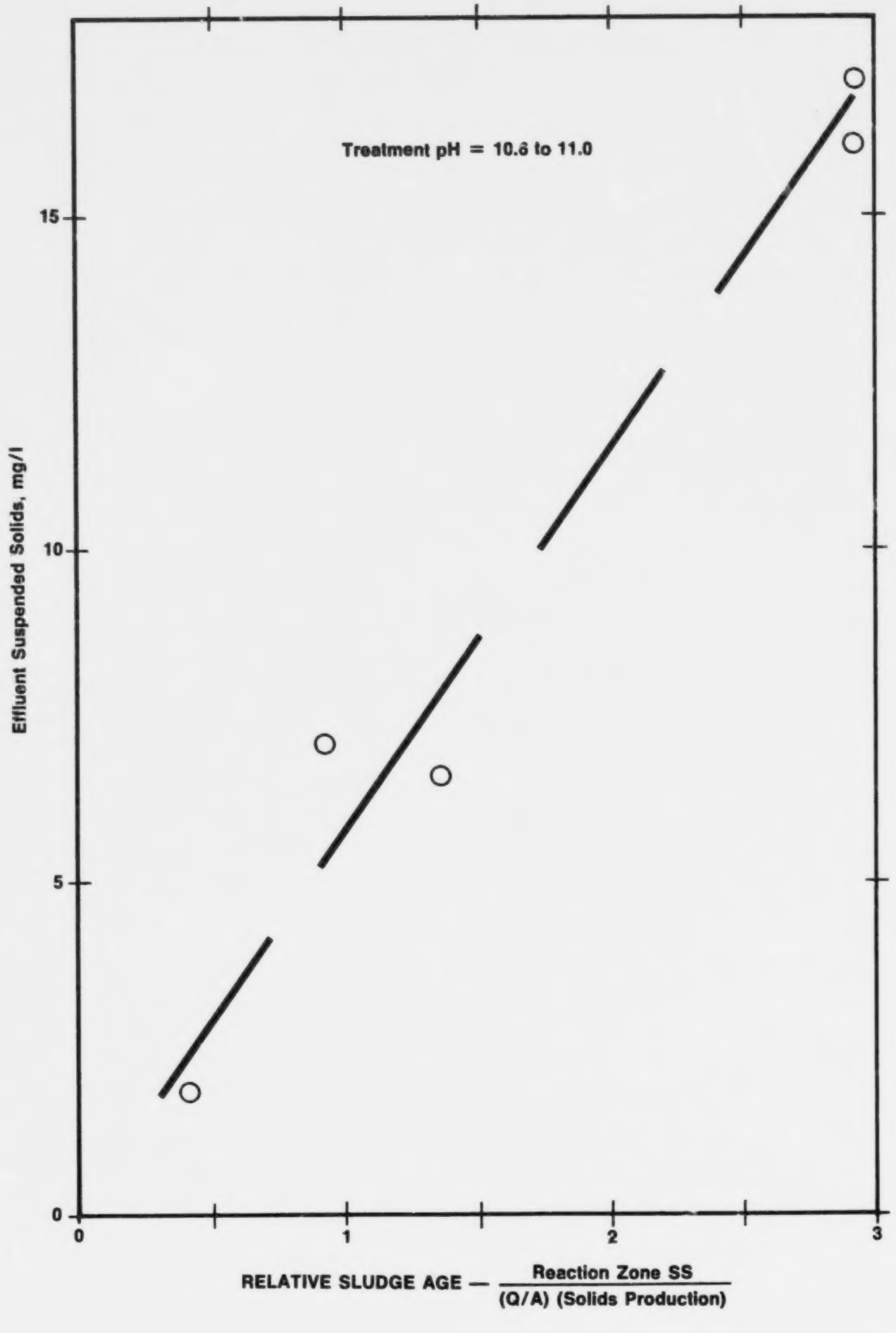


FIGURE 31: LIME TREATMENT: EFFECT OF SLUDGE AGE



Based on the data in Figure 31 and considering the operation conditions existing during pilot plant operation, it appears that reaction-zone suspended solids should be kept below about 4 to 6 g/ℓ. Effluent suspended solids would then be expected to average less than about 10 mg/ℓ.

The highest overflow rate evaluated over a sustained period of time was 1.3 gpm/sq ft, during Run L-5 at a pH of about 11.0. Just prior to Run L-5 overflow rates as high as 1.5 gpm/sq ft were tried, but control of the sludge level in the solids-contract unit proved difficult. Thus, for moderate treatment pH, a maximum sustained overflow rate of about 1.3 gpm/sq ft is suggested. The hydraulic capacity of the pilot plant was limited to an overflow rate of 1.5 gpm/sq ft. Consequently, an evaluation of the effect of polyelectrolyte flocculation aid on clarifier overflow rate was impossible.

Because of limited lime feeding facilities, the maximum hydraulic loading for high treatment pH (11.5) Runs could not be determined. During Runs L-10 and L-11 excellent clarification was obtained at overflow rates from 0.4 to 0.7 gpm/sq ft. As will be developed later, chemical sludge production and sludge properties were somewhat similar for moderate and high treatment pH values. It is, therefore, conjectured that the maximum overflow rate for high treatment pH would be similar to that experienced for moderate treatment pH's.

Data in Table 9 indicates a range of SCOD removals from 13 percent to -41 percent (i.e., an increase) across the chemical contactor. These data were quite variable and did not precisely correlate with operational parameters such as relative sludge age, treatment pH or hydraulic retention time at treatment pH. Since any increase in SCOD across the chemical treatment step will result in an increased demand for powdered carbon in the subsequent treatment step, future lime treatment studies should be designed to more precisely evaluate SCOD removal.

The hardness data in Table 9 indicate that some softening was accomplished at treatment pH's between 10.6 to 11.1. Above and below this pH range hardness was added to the wastewater. At all pH's tested calcium hardness was found to increase. Some 80 to 85 percent of the biocarbonate alkalinity was removed at moderate treatment pH values. When lime at dosages above 600 mg/ℓ was added to reach high treatment pH levels, considerable calcium increases occurred because $CO_3^=$ was not available to precipitate the additional Ca^{++} as $CaCO_3$. If high dissolved solids are objectionable, then recarbonation of high treatment pH effluent to pH 9.5 to 10.5 with CO_2 would be required to reduce the high calcium concentration. A batch test during Run L-11 indicated that recarbonation with CO_2, to a pH of 10.4, reduced soluble calcium from 1500 mg/ℓ to about 160 mg/ℓ Ca^{++} (as $CaCO_3$). This recarbonation resulted in some 11,000 lb of $CaCO_3$ sludge per MG of wastewater flow (i.e., 1320 mg/ℓ of solids precipitated).

Sludge Production

Table 10 presents a summary of observed chemical and sewage sludge production data. Chemical sludge production is seen to range from 551 to 1016 mg/ℓ at moderate treatment pH values and from 690 to 1289 mg/ℓ for high treatment pH values. The variability in the data prompted an attempt to correlate chemical sludge production with treatment parameters such as pH, lime dosage and alkalinity. The best correlation found is shown in Figure 32 which indicates that the pounds of chemical sludge produced per pound of lime fed decreased in the pH range 10.7 to 11.1. Jar test results presented in Figure 8, did not show the leveling off at high treatment pH's.

Because of the variability of raw wastewater alkalinity and actual treatment pH between Runs, no precise correlation between lime dosage required to attain a given pH and alkalinity was found.

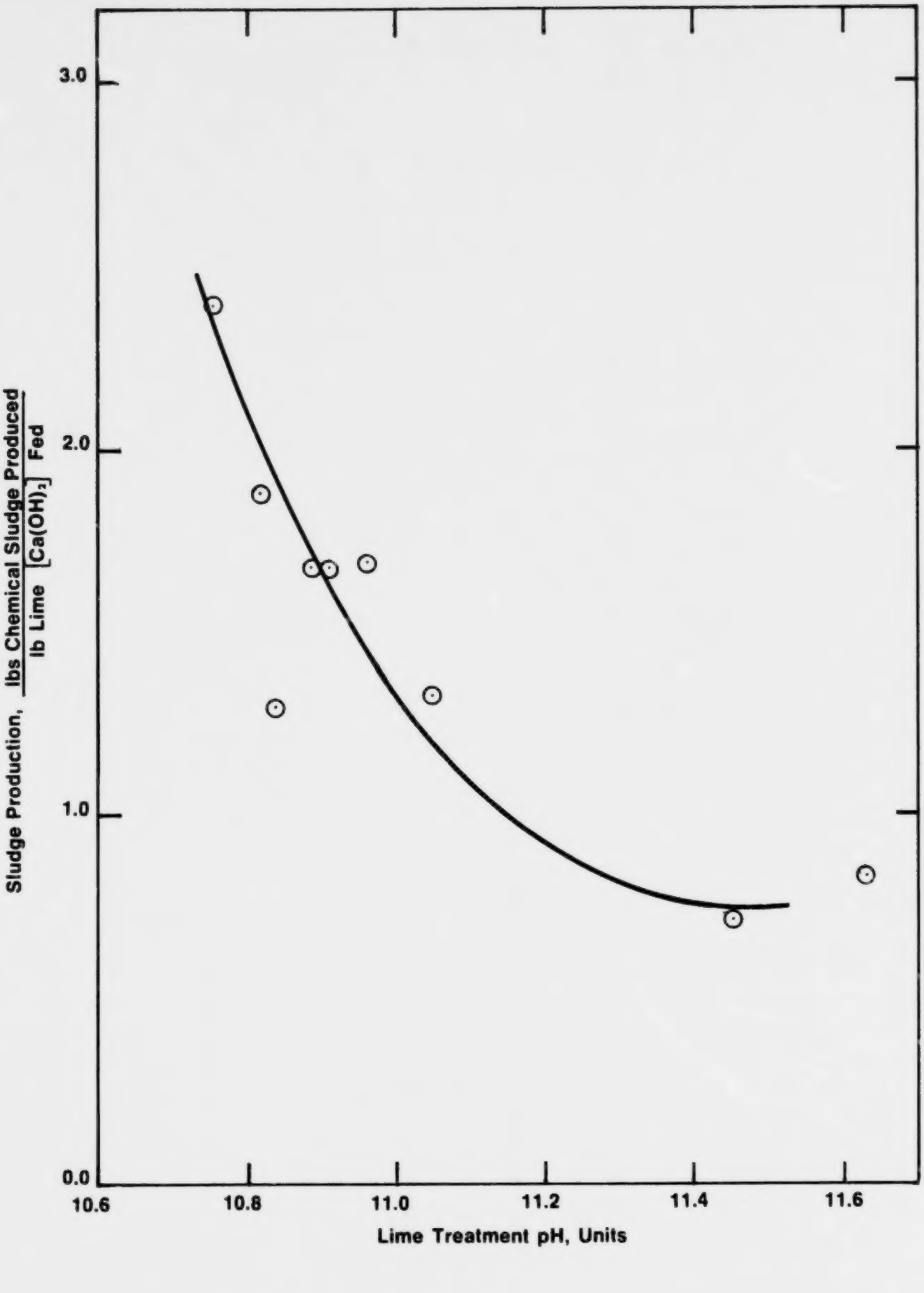
The following assumptions were made in an effort to relate chemical sludge production to changes in water quality parameters caused by lime treatment.

TABLE 10: LIME TREATMENT: SLUDGE PRODUCTION

Run #	3	4	5	6	7	8	9	10	11
Treatment pH	10.7	10.8	10.8	10.9	10.9	11.0	11.0	11.3-11.6	11.6
Blowdown SS, lb/MG[a]	9170	7010	5250	5840	8340	7340	7260	6300	12,100
Blowdown SS, mg/ℓ[a]	1100	842	630	700	1000	880	872	756	1450
Clarifier Effluent SS, mg/ℓ	16	8	13	12	7	17	2	6	7
Total Chemical and Sewage SS, mg/ℓ	1116	850	643	712	1007	897	874	762	1457
Raw Wastewater SS, mg/ℓ	100	170	92	~56	120	58	92	72	170
Chemical SS Produced, mg/ℓ	1016	680	551	656	887	839	782	690	1287
Lime Dosage, mg/ℓ $Ca(OH)_2$	425	365	420	390	525	490	585	800-1100	1530
lbs Chemical Solids/ lb $Ca(OH)_2$ Fed	2.39	1.86	1.31	1.68	1.68	1.70	1.33	0.73	0.48
% Chemical Solids	92	80	86	93	88	94	90	91	89

a - Based on Total Flow

FIGURE 32: LIME TREATMENT: SLUDGE PRODUCTION



1. The reduction of bicarbonate alkalinity results from precipitation of $CaCO_3$
2. The reduction in non-calcium hardness results from precipitation of $Mg(OH)_2$
3. The reduction in phosphorus results from precipitation of $Ca_5OH(PO_4)_3$

Based on these assumptions, computations indicated that about 470 mg/ℓ of $CaCO_3$ was produced at moderate treatment pH values (10.6 to 11.1). The computed chemical sludge distribution was found to be 85 percent $CaCO_3$, 10 percent $Mg(OH)_2$ and 5 percent $Ca_5OH(PO_4)_3$. A mass balance of calcium indicated that some 14 percent of the calcium in the sludge was present in a form other than $Ca_5OH(PO_4)_3$ or $CaCO_3$. This amount of calcium was equivalent to about 60 mg/ℓ of $Ca(OH)_2$. Computed chemical sludge values were 20 to 30 percent less than the observed chemical sludge production values in Table 10.

From Table 10 it is seen that 80 to 94 percent of the lime-sewage sludge produced was chemical sludge. From the above discussion, it would appear that moderate treatment pH, lime-sewage sludge consisted of the following approximate consitutents:

1. 10% sewage solids
2. 60% $CaCO_3$
3. 10% $Mg(OH)_2$ and $Ca_5OH(PO_4)_4$
4. 8% some other calcium precipitate
5. 12% unaccounted for chemical solids

From this estimate of the distribution of sludge constitutents, it is predictable that the large fraction of $CaCO_3$ should result in an easily thickened and dewatered lime-sewage sludge.

Sludge Thickening and Dewatering

Numerous laboratory gravity thickening and vacuum filter leaf tests were conducted on moderate and high treatment pH lime-sewage sludges. The effect of polyelectrolyte flocculation aid was also established.

Figure 33 shows the results of laboratory thickening tests. Each curve represents one sludge sample and shows the thickener loading (i.e., no scale-up factor) related to various thickened sludge concentrations. The data are somewhat erratic, reflecting variable sludge characteristics experienced over the eight month period during which the tests were conducted. It is apparent from the feed solids data in Figure 33 that the moderate treatment pH sludges tested were generally more concentrated than the high treatment pH sludges. This observation reflects the results shown in Table 9.

The lower-right extremity of the curves in Figure 33 represents the sludge thickness attained after 24-hours of gravity thickening and is defined as the maximum possible or ultimate concentration. A plot of maximum thickened concentration versus initial or feed solids concentration is presented in Figure 34. It appears that the more concentrated the chemical clarifier underflow (initial solids) the more concentrated are the settled solids after 24-hours of settling.

One approach to using the thickening data entails drawing an operating line through each set of curves. For the data presented, high and moderate treatment pH sludges are separately considered. Generally, higher vacuum filter yields and drier filter cakes are obtained with higher feed solids concentration (i.e., thickener underflow solids concentration). Since vacuum dewatering costs are considerably higher than gravity thickening costs, the operating lines in Figure 33 are drawn at reasonably high underflow solids concentration values. If the thickener solids loading and underflow concentration at the intersection of each thickening curve with the operating line is plotted versus thickener feed solids, a suggested design curve can be drawn.

Figure 35 presents such a plot of suggested design and thickening data for the high and moderate treatment pH sludges. A scale-up factor of 0.80 was applied to the observed thickener solids loading values taken from

FIGURE 33: LIME TREATMENT: SLUDGE THICKENING

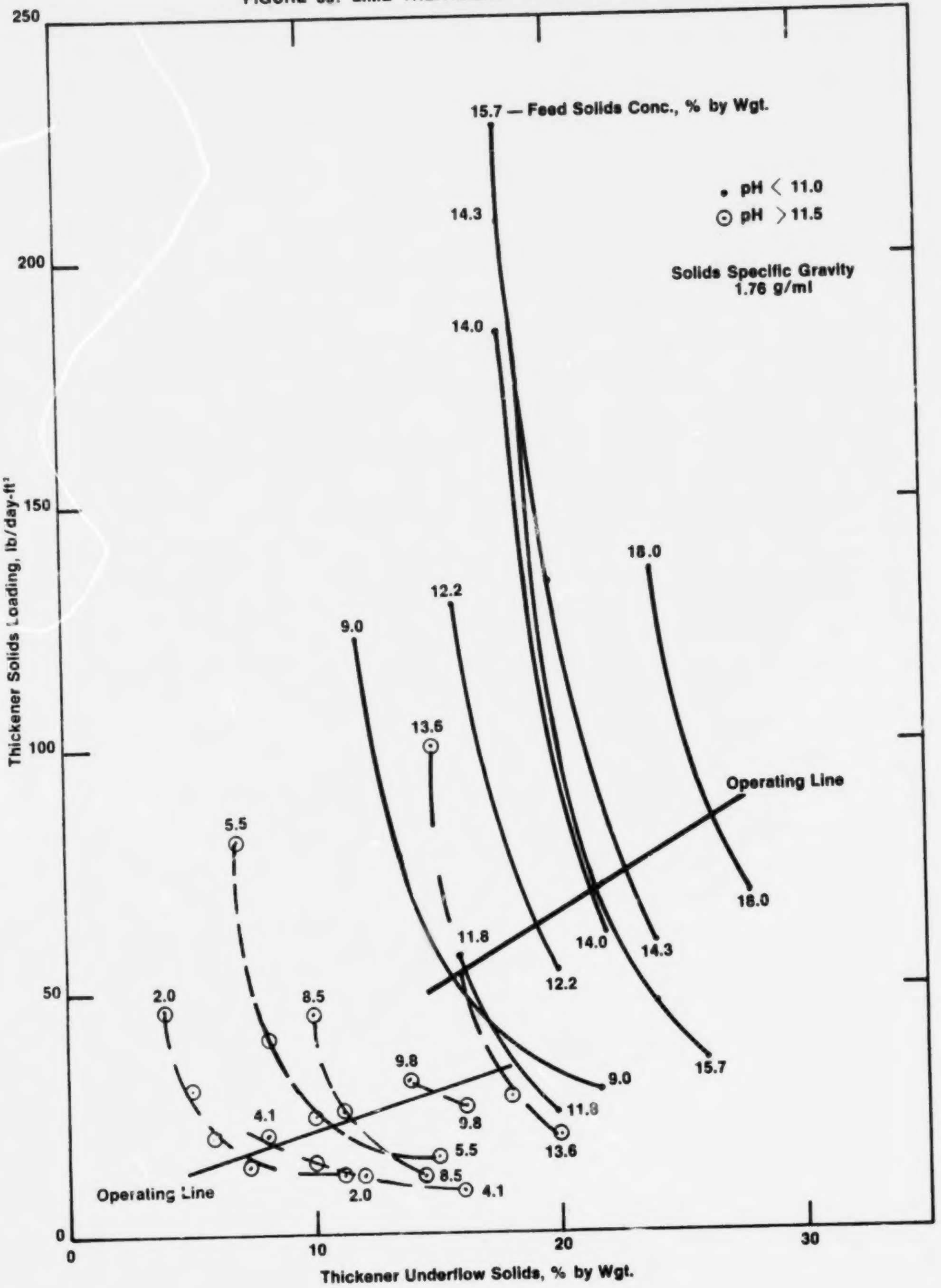


FIGURE 34: LIME TREATMENT: SLUDGE THICKENING

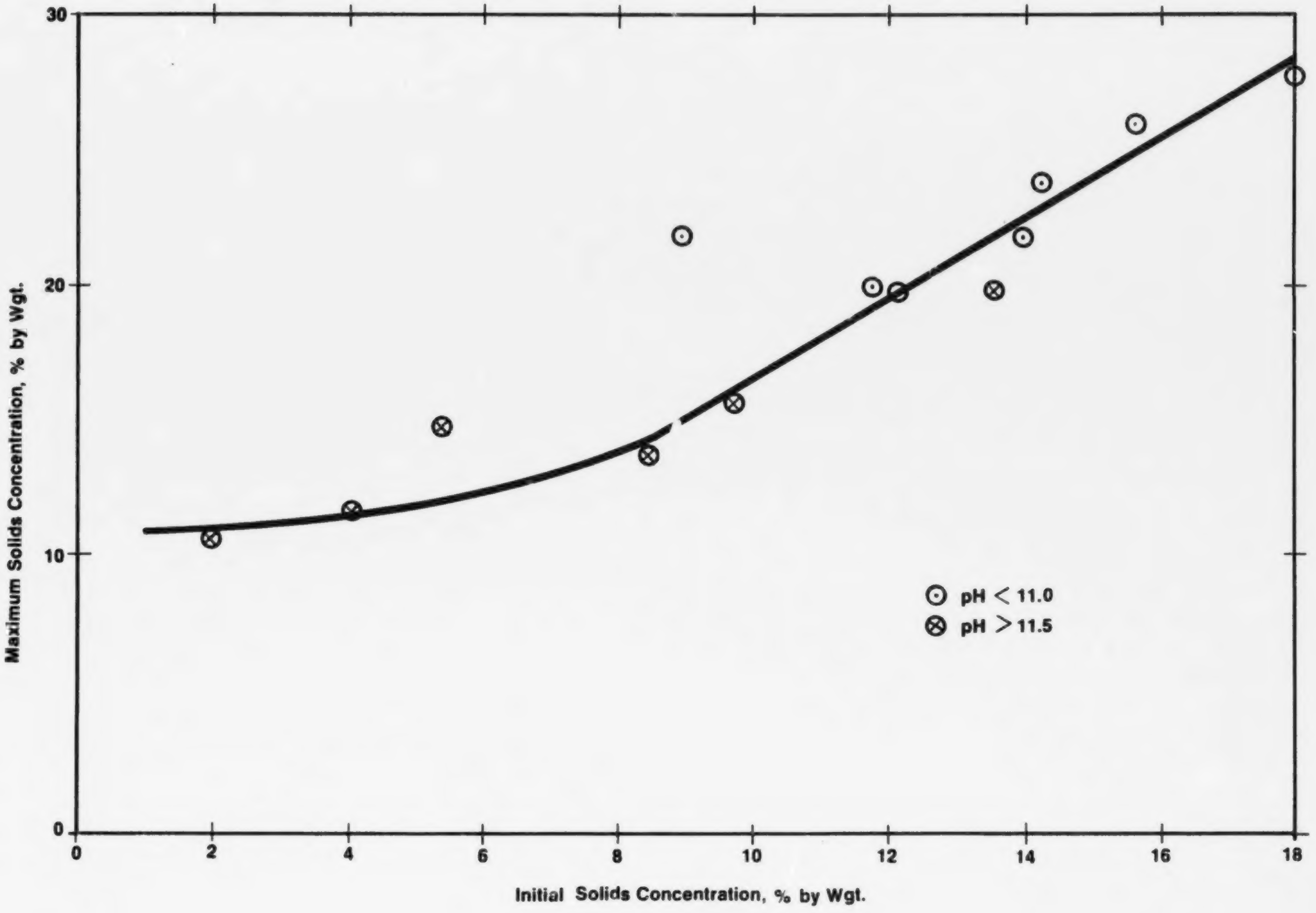


FIGURE 35: LIME TREATMENT: SLUDGE THICKENING

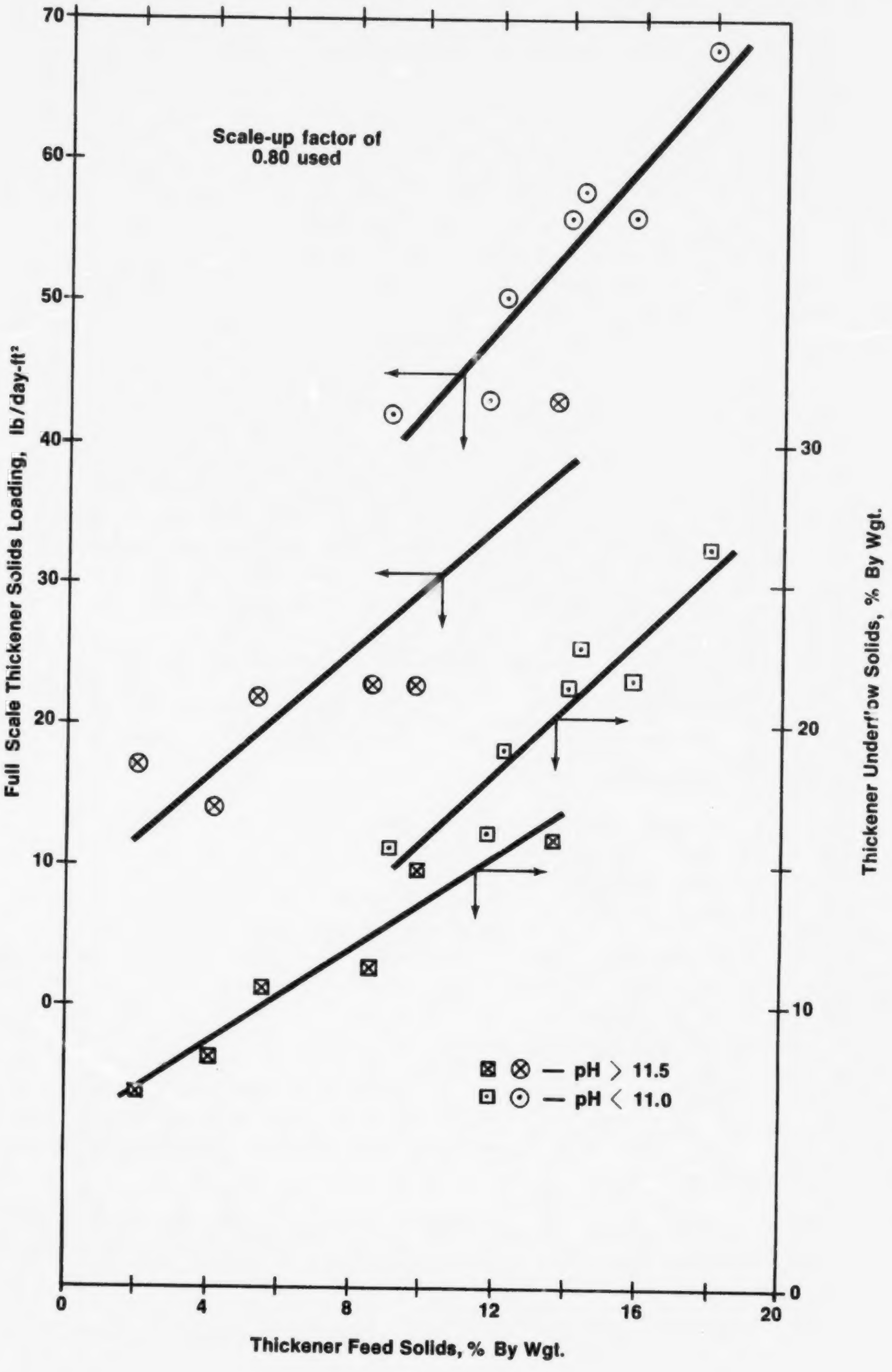


Figure 33. The linear regression lines shown in Figure 35 are for the two separate treatment pH ranges. A linear regression of all treatment pH values of thickener solids loading and underflow solids data on feed solids would result in correlation coefficients of 0.90 and 0.98 respectively. This observation would imply that the moderate and high treatment pH sludges had similar thickening properties. This implication is possibly dependent, however, on the choice of the location of the operating lines drawn in Figure 33 and, hence, is not a valid general conclusion.

The effect of polyelectrolyte flocculation aid on thickening results is indicated in Figure 36 for a sample of high treatment pH sludge. The data shows that the lower the feed solids concentration, the greater the effect on thickener solids loading. The cost of polyelectrolyte probably cannot be justified on the basis of reduced thickener size. Polyelectrolyte use on a part-time basis may be justified. For example, consider an installation designed for thickening 6 percent feed solids to 12 percent, at 21 lb/day-sq ft, without polyelectrolyte. If for some operational reason, the feed solids dropped to 3 percent the thickener would still produce 12 percent underflow if about 0.04 lb polyelectrolyte per lb dry sludge solids were added.

Numerous vacuum filter leaf tests were conducted on both high and moderate treatment pH sludges. A synthetic filter media, Polypropylene-852F, was found to result in good filtrate clarity of less than 200-300 mg/ℓ suspended solids and would discharge the cake easily. The filter leaf tests results are presented in Figure 37. Maximum predicted vacuum filter yields for the operating conditions indicated are shown. The 33 percent drum submergence and 20 inches of Hg vacuum level are acceptable design values. The results in Figure 37 indicate that treatment pH and feed solids significantly affect the vacuum filter yield. The linear regression lines shown have correlation coefficients of 0.90 and 0.96 for high and moderate treatment pH values respectively.

Several tests were conducted with anionic polyelectrolyte flocculation aid (Dow Chemical Co. AP-30) at dosages ranging from 0.2 to 0.6 lb/ton dry solids. Vacuum filter yields were found to increase from 30 to 70 percent above those shown in Figure 37 for several feed solids concentration.

FIGURE 36: LIME TREATMENT: SLUDGE THICKENING

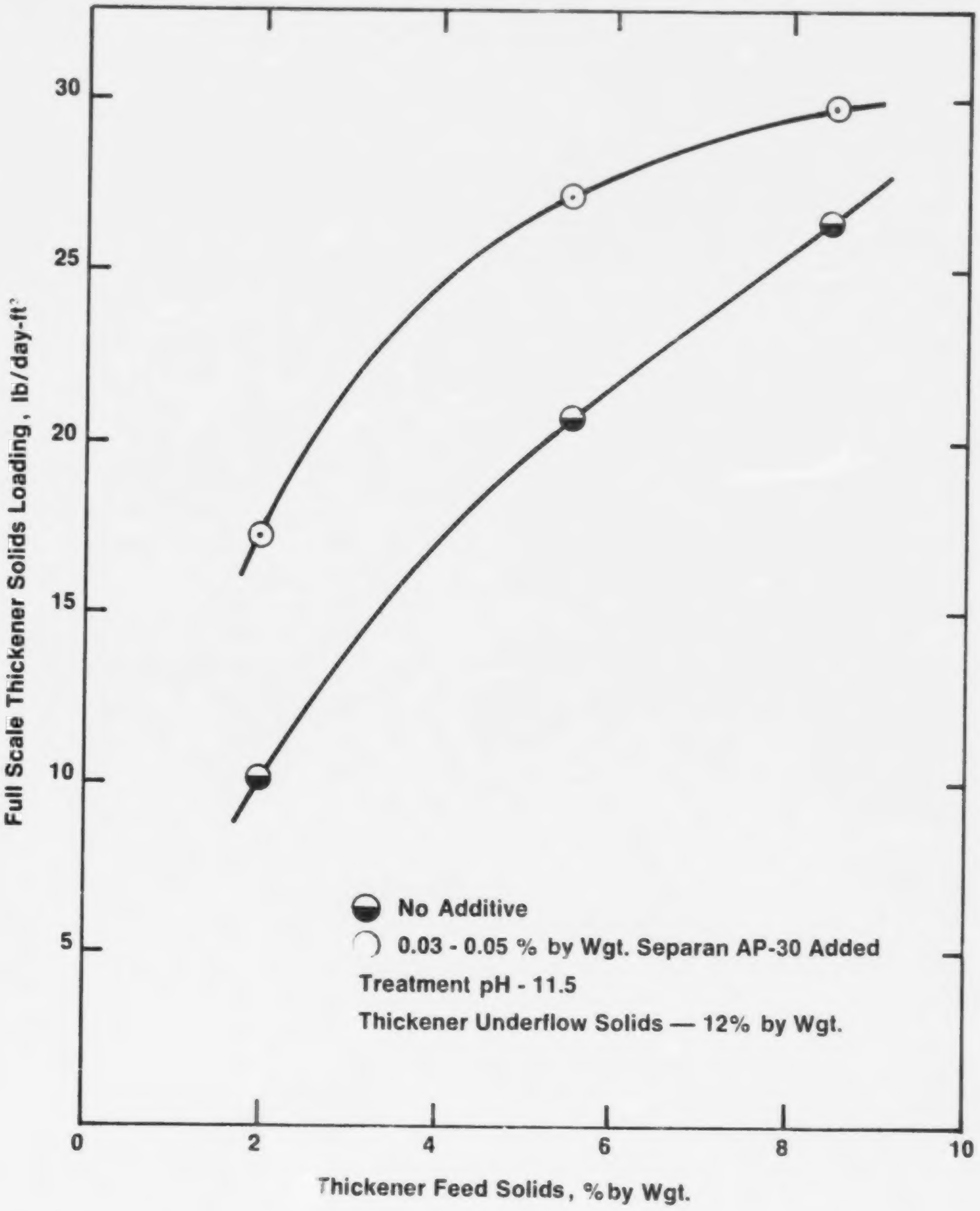
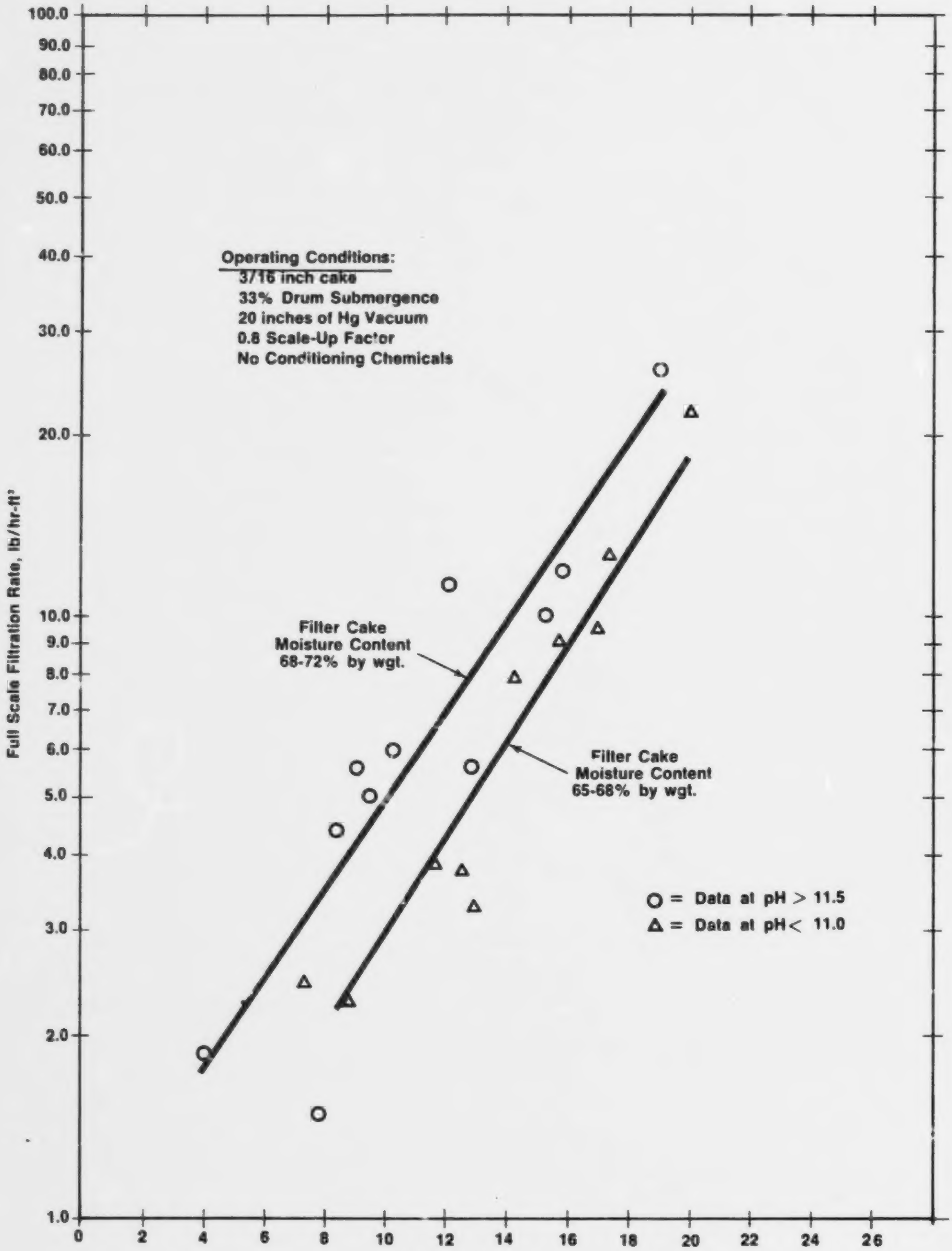


FIGURE 37: LIME TREATMENT SLUDGE DEWATERING



The utility of the lime-sewage sludge thickening and dewatering data presented can best be illustrated by the following example which indicates suggested design criteria and results for lime-sewage sluge produced in this study. Table 11 is a summary of the following example:

Chemical clarifier underflow is taken as the weighted average of all such data presented in Table 9. From Figure 35, thickener solids loading and underflow concentration is determined. For the observed thickener underflow concentration (vacuum filter feed solids) a filter yield is determined from Figure 37. For the vacuum filter operating conditions specified in Figure 37, the following relationships would exist:

$$CT = \frac{FT}{0.33} \tag{2}$$

$$\Theta_D = 0.50\ CT \tag{3}$$

$$W = \text{Yield} \times \frac{CT}{60} \tag{4}$$

Where: CT = drum cycle time, min

FT = cake form time, min

Θ_D = cake dry time, min

W = cake weight, lb/sq ft

Yield = full scale filtration rate, lb/hr-sq ft

Figure 39 presents the observed relationship between filter cake moisture content and a modified correlating factor (Θ_D/W) for the treatment pH and filter feed solids conditions indicated. Determination of filter cake moisture content presented in Table 11 was as follows (example is for moderate treatment pH):

1. $CT = \frac{0.6}{0.33} = 1.8$ min
2. $\Theta_D = (0.5)(1.8) = 0.9$ min

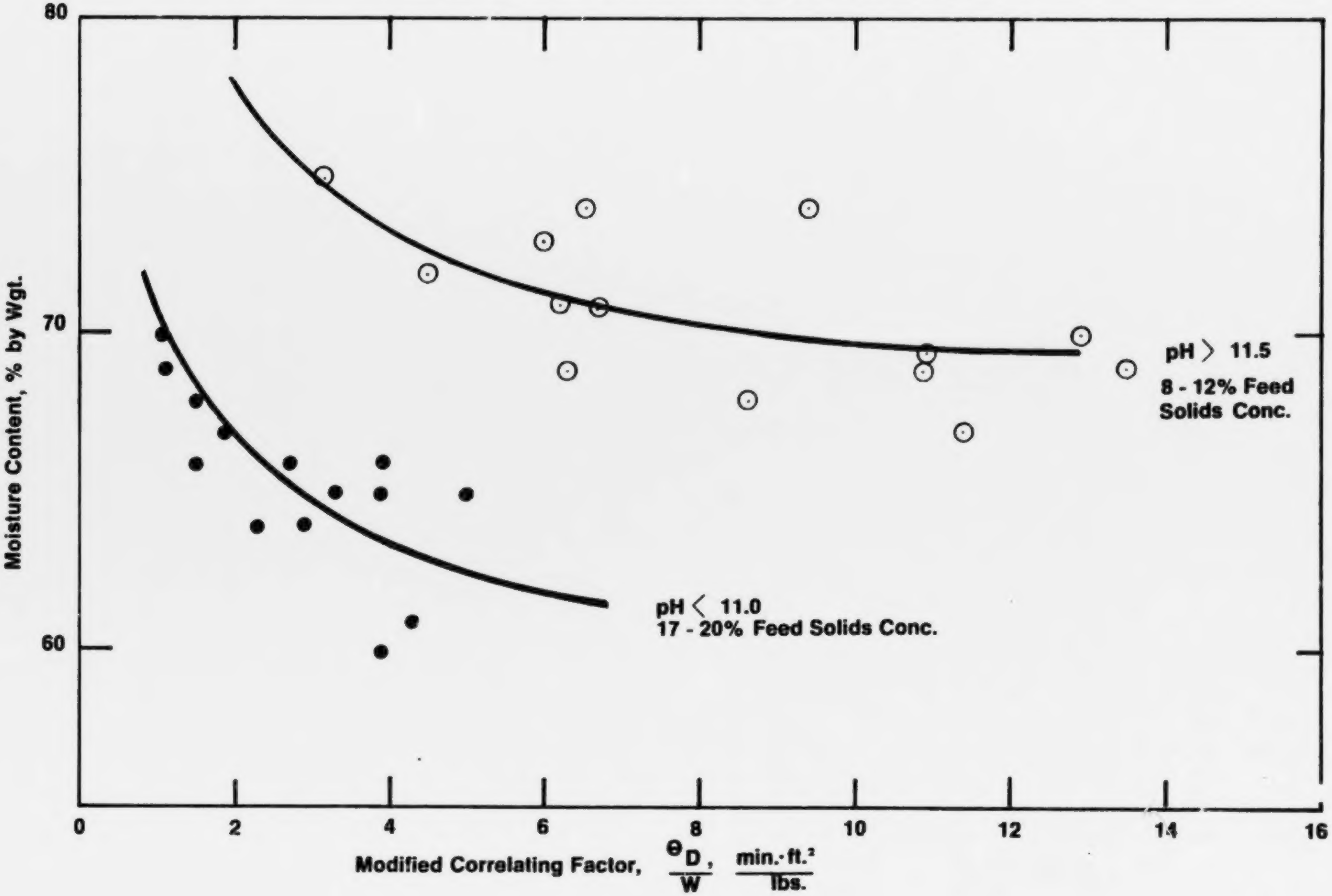
   $W = \frac{(12.8)(1.8)}{60} = 0.38$ lb/sq ft
3. $(0.8)\ \frac{\Theta_D}{W} = \frac{(0.9)(0.8)}{0.38} = 1.9$ min-sq ft/lb

TABLE 11: SUGGESTED DESIGN PARAMETERS FOR LIME-SEWAGE SLUDGE THICKENING AND DEWATERING

	Treatment pH	
Item	Moderate <11.0	High >11.5
Chemical Clarifier underflow concentration (weighted avg., of all data) % by wgt	11.5	4.8
Predicted thickener underflow solids[a] - % by wgt	17.9	9.4
Predicted thickener solids loading[a] - lb/day-sq ft	47	18
Predicted vacuum filter yield[b] - lb/hr-sq ft	12.8	4.5
Required filter cake form time - min (from leaf test results)	0.6	1.37
Modified correlating factor-min-sq ft/lb (includes 0.8 scale-up factor)	1.9	5.4
Filter Cake moisture[c] - % by wgt	67	72

a = from Figure 35
b = from Figure 37
c = from Figure 38

FIGURE 38: LIME TREATMENT: SLUDGE DEWATERING



From Figure 39 a moisture content of 67 percent is indicated at $\Theta_D/W = 1.9$.

Summary of Chemical Treatment Results

As predicted by laboratory jar tests, all three chemicals tests were effective in precipitating phosphorus and coagulating suspended solids. The results indicate that Salt Lake City raw municipal wastewater could be chemically treatment to produce an effluent of about 10 JTU Turbidity, 10-25 mg/ℓ suspended solids and 1 mg/ℓ phosphorus. The raw wastewater characteristics encountered varied widely; however, it could be generally classified as a moderately weak municipal wastewater. It is conjectured that similar treatment effectiveness could be obtained when treated a normal to moderately strong municipal wastewater.

To facilitate comparison of the three treatment chemicals studied, a summary is presented in Table 12. The chemical dosages and chemical contactor operating conditions presented should result in production of the above mentioned effluent quality. Though use of polyelectrolyte flocculation aid is not indicated, it is strongly recommended that facilities be provided for feeding polyelectrolyte to all three unit operations of clarification, thickening and vacuum dewatering. Doing so will provide reserve treatment capacity, thus enhancing performance reliability.

Because of the limited number of $FeCl_3$ and alum sludge thickening and filtration tests conducted, the data presented for these sludges should be considered preliminary. Additional tests are recommended.

It can be noted that high pH lime treatment results are not presented in Table 12. Any benefits achieved by high treatment pH, compared to moderate treatment pH, were offset by significantly increased lime requirements, increased sludge production and poorer sludge thickening characteristics. Actually, the high pH lime-sewage sludge thickening properties were significantly poorer than for moderate pH sludges.

The ultimate choice of which type of chemical should be used must be based on an economic analysis of total treatment costs. The design conditions presented in Table 12 will be used later in this report as the basis for such an economic analysis for treating Salt Lake City wastewater.

TABLE 12: SUMMARY OF SUGGESTED CHEMICAL TREATMENT PROCESS DESIGN PARAMETERS FOR SALT LAKE MUNICIPAL WASTEWATER

Treatment Chemical:	$FeCl_3$	Alum	Hydrated Lime
Chemical Dosage, mg/ℓ	120	140	400-500
Acid (H_2SO_4) Dosage, mg/ℓ	0	0	330
Chemical-Contactor:			
Peak Overflow Rate, gpm/sq ft	0.5	0.4	1.3
Reaction Zone Solids, % by wgt	0.2	0.1	0.5
Underflow Solids, % by wgt	1.3	0.5-1.0	11.5
Sludge Production:			
Chemical-Sewage Sludge, mg/ℓ[a]	180	145	850
% Chemical Sludge	52	40	90
lb Chem. Sludge/lb Chem. Fed	0.78	0.41	1.7
Gravity Thickening:			
Solids Loading, lb/day-sq ft	10	5-8	47
Underflow Concentration, % by wgt	4.0	2.5	18
Vacuum Filtration:			
Yield, lb/hr-sq ft	1.0[b]	0.6-1.0[b]	12.5
Filter Cake Moisture, % by wgt	82	82	67

---

a = based on total plant flow
b = with conditioning chemicals: 20-25% $Ca(OH)_2$ or 0.5% polymer.

The problem of "post precipitation" of iron compounds remains imprecisely defined and without a proven solution. If the problem is ultimately defined as being the reduction and subsequent oxidation of soluble iron, then a possible solution can be suggested. In the PAC-PCT process evaluated, an oxidation operation could be accomplished after carbon contacting and prior to granular media filtration. Aeration at this point should effectively convert Fe^{+2} to Fe^{+3}, with precipitation of the latter as a filterable hydrated ferric oxide. Oxygen requirements and aeration contact time and method requirements will have to be determined before this approach can be accepted.

POWDERED CARBON TREATMENT STEP

The major purpose of the carbon treatment tests was to determine the effect of carbon dosage and number of contact stages (1 or 2) on carbon contactor operation and removal of soluble organics from chemically treated and gravity clarified raw wastewater. Results of spent carbon handling, concentration and regeneration are presented and discussed later in this report (see page 152).

Based on laboratory equilibrium adsorption isotherm test results, it was concluded that the type of chemical used for pretreatment did not significantly affect the adsorption of SCOD by carbon. Thus, the pilot plant carbon treatment Runs were scheduled independently of the chemical treatment Runs.

A summary of operating conditions and treatment results is presented in Tables 13 and 14. Chronologically, Runs C-14 and C-13 were conducted first, followed by Runs C-2, C-1 and then C-3 through C-12. Certain single-stage contacting Runs were conducted simultaneously by operation of the two carbon contactors in parallel (for example, Runs C-9 and C-10; C-11 and C-12). The chemical pretreatment Run numbers shown refer to Tables 2, 5 and 9. It should be noted that the carbon and chemical treatment Run periods generally do not precisely coincide, thus chemical stage effluent results differ slightly between Tables 13 and 14 and Tables 2, 5 and 9.

The approximate solids retention time (SRT) of the carbon in the contactors was computed as follows:

$$\text{SRT} = \frac{(\text{Reaction Zone Solids})(\text{Volume of Carbon Slurry})}{(\text{Flow Rate})(\text{Carbon Dosage})}$$

TABLE 13: SUMMARY OF TWO-STAGE COUNTER-CURRENT CARBON TREATMENT RESULTS

Run #		C-1				C-2		
Length of Run, Days		12				34		
OPERATING CONDITIONS:								
Chemical Pretreatment, (Run #)		L-3				L-8		
Hydraulic Loading, gpm/sq ft (1st stage/2nd stage/GMF)		0.76/0.73/4.2				0.69/0.69/3.1		
Carbon Dosage, mg/ℓ		325				150		
Polyelectrolyte Dosage, mg/ℓ (1st stage/2nd stage/GMF)		0/0/0				0.3/0.3/0.3		
Reaction Zone SS, g /ℓ (1st stage/2nd stage)		17/19				9/8		
Blowdown Volume, % of Flow		0.9				--		
Approximate SRT, days (1st stage/2nd stage)		2.0/2.2				2.4/2.2		
PROCESS TREATMENT RESULTS:								
Effluent From	Chem. Stage	1st Stage	2nd Stage	GMF	Chem. Stage	1st Stage	2nd Stage	GMF
Suspended Solids, mg/ℓ	14	12	7	3	32	38	17	12
Turbidity, JTU	6	6	2	1	10	6	4	4
% Transmittance	--	--	--	--	--	--	--	--
SCOD, mg/ℓ	45	12	11	12	42	23	18	19
STOC, mg/ℓ	--	--	--	--	22	15	12	12
BOD_5, mg/ℓ	--	--	--	--	47	31	16	13
pH, units	7.1	--	--	7.5	8.3	--	--	8.0
Temperature, °C	22	--	--	--	22	--	--	--

TABLE 13 (cont.): SUMMARY OF TWO-STAGE COUNTER-CURRENT CARBON TREATMENT RESULTS

Run #	C-3				C-4			
Length of Run, Days	23				8			
OPERATING CONDITIONS:								
Chemical Pretreatment, (Run #)	L-10				L-10			
Hydraulic Loading, gpm/sq ft (1st stage/2nd stage/GMF)	0.56/0.52/1.4				0.79/0.79/4.7			
Carbon Dosage, mg/ℓ	110				75			
Polyelectrolyte Dosage, mg/ℓ (1st stage/2nd stage/GMF)	0/0/0				0/0/0			
Reaction Zone SS, g /ℓ (1st stage/2nd stage)	6/10				2/2			
Blowdown Volume, % of Flow	--				--			
Approximate SRT, Days (1st stage/2nd stage)	2.7/4.5				1.0/1.0			
PROCESS TREATMENT RESULTS:								
Effluent From	Chem. Stage	1st Stage	2nd Stage	GMF	Chem. Stage	1st Stage	2nd Stage	GMF
Suspended Solids, mg/ℓ	8	12	6	2	7	16	12	3
Turbidity, JTU	7	3	2	3	6	4	3	1
% Transmittance	--	--	--	--	--	--	--	--
SCOD, mg/ℓ	63	30	22	22	66	39	32	36
STOC, mg/ℓ	--	--	--	--	--	--	--	--
BOD_5, mg/ℓ	49	18	12	13	--	--	--	--
pH, units	6.0	--	--	6.3	8.2	--	--	8.3
Temperature, °C	21	--	--	--	19	--	--	--

TABLE 13 (cont.): SUMMARY OF TWO-STAGE COUNTER-CURRENT CARBON TREATMENT RESULTS

Run #	C-8				C-14			
Length of Run, Days	29				16			
OPERATING CONDITIONS:								
Chemical Pretreatment, (Run #)	L-7,L-9				F-2			
Hydraulic Loading, gpm/sq ft (1st stage/2nd stage/GMF)	0.45/0.40/1.6				0.75/0.73/1-2			
Carbon Dosage, mg/ℓ	150				600			
Polyelectrolyte Dosage, mg/ℓ (1st stage/2nd stage/GMF)	0/0/0				0.7/0.8/0			
Reaction Zone SS, g /ℓ (1st stage/2nd stage)	14/17				~15/~21			
Approximate SRT, Days (1st stage/2nd stage)	6.1/7.4				0.9/1.3			
PROCESS TREATMENT RESULTS:								
Effluent From	Chem. Stage	1st Stage	2nd Stage	GMF	Chem. Stage	1st Stage	2nd Stage	GMF
Suspended Solids, mg/ℓ	6	16	3	1.3	13	28	7.5	2.7
Turbidity, JTU	4	7	5	1	16	13	11	5
% Transmittance	98	82	93	--	--	--	--	--
SCOD, mg/ℓ[a]	45	18	11	4.7	37	15	15	10
STOC, mg/ℓ[a]	24	11	9	6	28	15	12	13
BOD_5, mg/ℓ	--	--	--	--	--	--	--	--
pH, Units	7.7	7.4	7.4	7.5	6.9	--	--	--
Temperature, °C	15-18	--	--	--	--	--	--	--

---

[a] - For Run C-14 only Total COD and TOC shown

TABLE 14: SUMMARY OF SINGLE-STAGE CARBON TREATMENT RESULTS

Run #	C-5		C-6			C-7		
Length of Run, Days	21		14			16		
OPERATING CONDITIONS:								
Chemical Pretreatment, (Run #)	L-6, L-1		L-6, A-1			A-1, A-2		
Hydraulic Loading, gpm/sq ft (Carbon Contactor/GMF)	0.53/0		0.53/3.7			0.68/2.6		
Carbon Dosage, mg/ℓ	95		300			105		
Reaction Zone SS, g /ℓ	8.3		3.4			3.2		
Blowdown Volume, % of flow	0.23		0.65			0.59		
Approximate SRT, Days	4.6		0.6			1.2		
PROCESS TREATMENT RESULTS:								
Effluent From	Chem. Stage	Carbon Cont.	Chem. Stage	Carbon Cont.	GMF	Chem. Stage	Carbon Cont.	GMF
Suspended Solids, mg/ℓ	14	24	21	14	3	27	10	2
Turbidity, JTU	9	7	15	7	2	11	6	2
% Transmittance	--	--	--	--	--	--	--	--
SCOD, mg/ℓ	58	24	36	10	18	28	16	16
STOC, mg/ℓ	28	10	25	13	12	16	13	11
BOD_5, mg/ℓ	58	21	35	23	16	39	8	10
pH, Unit	7.2	7.4	7.6	--	7.5	7.0	--	6.8
Temperature, °C	18	--	--	--	--	17	--	--

TABLE 14 (cont.): SUMMARY OF SINGLE-STAGE CARBON TREATMENT RESULTS

Run #	C-9			C-10		C-11		
Length of Run, Days	24			24		20		
OPERATING CONDITIONS:								
Chemical Pretreatment, (Run #)	L-11			L-11		L-4		
Hydraulic Loading, gpm/sq ft (Carbon Contactor/GMF)	0.30/1.2			0.30/--		0.75/2-5		
Carbon Dosage, mg/ℓ	146			146		345		
Reaction Zone SS, g /ℓ	9.5			11.5		--		
Blowdown Volume, % of flow	--			--		--		
Approximate SRT, Days	5.9			7.2		2.2		
PROCESS TREATMENT RESULTS:								
Effluent From	Chem. Stage	Carbon Cont.	GMF	Chem. Stage	Carbon Cont.	Chem. Stage	Carbon Cont.	GMF
Suspended Solids, mg/ℓ	6	30	2	6	30	7.7	53	1.8
Turbidity, JTU	4	6	2	4	9	5	8	2
% Transmittance	98	91	99	98	81	97	76	99
SCOD, mg/ℓ	35	24	23[a]	35	25	41	6.4	3.6
STOC, mg/ℓ	--	--	--	---	--	--	--	--
BOD5, mg/ℓ	34	13	13	34	17	--	--	--
pH, Unit	7.6	2-11	8.0	7.6	2-11	--	--	--
Temperature, °C	15	--	--	15	--	22	--	--
Copper, mg/ℓ	--	--	--	--	2.4	--	--	--

---

a - Total COD

TABLE 14 (cont.): SUMMARY OF SINGLE-STAGE CARBON TREATMENT RESULTS

Run #	C-12		C-13	
Length of Run, Days	20		13	
OPERATING CONDITIONS:				
Chemical Pretreatment, (Run #)	L-4		F-3	
Hydraulic Loading, gpm/sq ft (Carbon Contactor/GMF)	0.75/--		0.75/--	
Carbon Dosage, mg/ℓ	317		0	
Reaction Zone SS, g /ℓ	--		6	
Blowdown Volume, % of flow	--		0	
Approximate SRT, Days	1.6		∞	
PROCESS TREATMENT RESULTS:				
Effluent From	Chem. Stage	Carbon Cont.	Chem. Stage	Carbon Cont.
Suspended Solids, mg/ℓ	7.7	20	18	18
Turbidity, JTU	5	5	31	32
% Transmittance	97	90	--	--
SCOD, mg/ℓ[b]	41	13.2	55	38
TOC, mg/ℓ[b]	--	--	21	15
pH, Unit	7.6	7.8	--	--
Temperature, °C	22	--	--	--

---

b - For Run C-13, Total COD and TOC shown

The numerator is an approximation of the inventory of carbon (pounds) within the solids-contact units and the denominator is the carbon fed (lb/day). The carbon inventory, so calculated, was found to be within ±25 percent of that determined by a more comprehensive procedure to be presented later.

STOC data is presented for only a few Runs. Considerable difficulty was experienced maintaining the TOC analyzer in an operable condition. The need for frequent repairs and delayed delivery of parts were common experiences. The BOD_5 data presented in Tables 13 and 14 are for only 5 to 20 percent of the number of samples for SCOD data.

Operation of Carbon Contactors

As indicated in Appendix C (Operation of Solids-Contact Units), the principal operational feature of the carbon contactors was maintenance and control of copious amounts of powdered carbon within the treatment units. The 5-minute slurry settling test was routinely used to monitor reaction zone carbon concentration (see Appendix C).

The volume of carbon slurry within the carbon contactors was estimated by observation of slurry pool depth. The approximate location of slurry pool depth was determined by withdrawing samples from clarification zone sample taps, which were located at depths of 13, 30 and 58 inches from the bottom of the 126 inch high clarifier tank wall. The lower extremity of the reaction zone skirt was located at a depth of 22 inches. Normal operation entailed maintaining the carbon slurry pool just above the middle sample tap. If the 5-minute settling test of a sample taken from the top sample tap was more than 5-10 percent, the carbon blowdown rate was increased. This caused the slurry pool to fall below the middle sample tap within one-half to two hours. This operational procedure resulted in a 0 to 2 1/2 foot deep carbon slurry pool (i.e., sludge blanket) existing in the variable area clarification zone.

Carbon inventory within the contactors was determined on three occasions. Samples from three taps in the clarification zone and a single tap in the reaction zone were analyzed for carbon slurry concentration. Based on the geometry of the units, volumes were associated with each sample tap and a total weight of carbon computed. An example is shown in Table 15. The results of all carbon inventories indicated that multiplication of reaction zone suspended solids by total volume of concentrated carbon slurry within the units would estimate carbon inventory within ±25 percent. This fact is understandable since the reaction zone volume constitutes about 2/3 of the total slurry volume and the variation of suspended solids between sample taps was usually less than ±20 percent.

The uniform distribution of carbon and the "total solids circulation" feature of the solids-contact units provided the basis for a good biological reactor. During pilot plant start-up and shakedown, before solids inventory control was attempted, significant odors were noted at the surface of the carbon contactors (especially above the reaction zone). The odors were typical of anaerobic decomposition with occasional sulfide odor occurring. Addition of 1-2 SCFM of air per 10 gpm of flow into the reaction zone significantly reduced the anaerobic odors but did not eliminate them.

During Run C-14, a minimal odor level existed. The high carbon dosage (600 mg/ℓ) prompted an evaluation of carbon residence time (SRT). The rationale applied was that adsorption of biodegradable organics at the carbon surface could enhance bioactivity and that a biomass could be located at the external surface of the carbon particles. Thus, carbon SRT may approximate biomass SRT. Anaerobic biological kinetic studies have indicated that methane fermentation systems cannot be maintained at SRT's less than about 5-7 days and sulphate reduction systems at SRT's less than about 2-4 days[27,28].

Based on the above rationale, and the fact that during Run C-14 carbon SRT was less than about 1.5 to 2.5 days, it was considered desirable to attempt to maintain carbon SRT to less than about 2-3 days in each carbon contactor for odor control.

Data in Tables 13 and 14 show that attempts to control SRT were not always successful. Indeed, during Run C-8 SRT's of approximately 6-7 days actually occurred.

TABLE 15: DETERMINATION OF SOLIDS INVENTORY WITHIN A CARBON CONTACTOR *

Sample Tap Location	Solids Concentration g /ℓ	Volume 1000 liters	Solids Kg
Reaction Zone	18.1	7.0	127
Bottom Tap[a]	17.1	2.5	44
Middle Tap[a]	14.0	1.8	25
Top Tap[a]	nil	--	--
TOTALS		11.3	196

* Run C-14, first-stage contactor

Estimated Solids = Reaction zone solids x slurry volume
= 18.1 g /ℓ (11,300 ℓ) 10^{-3} Kg /g
= 205 Kg ≃ 196 Kg

a = sample taps at 13, 30 and 58 inches from the tank bottom and located at the wall

However, no significant odor problems were experienced. Possibly the low temperature of 15-18°C and variation in SRT due to operation (e.g., variations in carbon sludge blowdown rate to control carbon slurry pool height) minimized the chance of maintaining a prolific anaerobic system. It will be indicated later in this section that substantial biological removal of soluble organics (SCOD) probably occurred within the carbon contactors. Thus, elimination of anaerobic biological activity would not be considered desirable. Obviously, future studies should be designed to quantify odor problems (e.g., carbon contactor off-gas analysis) and evaluate the effect of plant operation (e.g., SRT) on anaerobic activity.

An attempt to quantify sulfide production was made during several Runs. Table 16 shows grab sample sulfide profiles through the pilot plant. It is apparent that excessive net increase of sulfide was not found across any of the PCT unit operations.

Clarification Effectiveness of Carbon Contactors

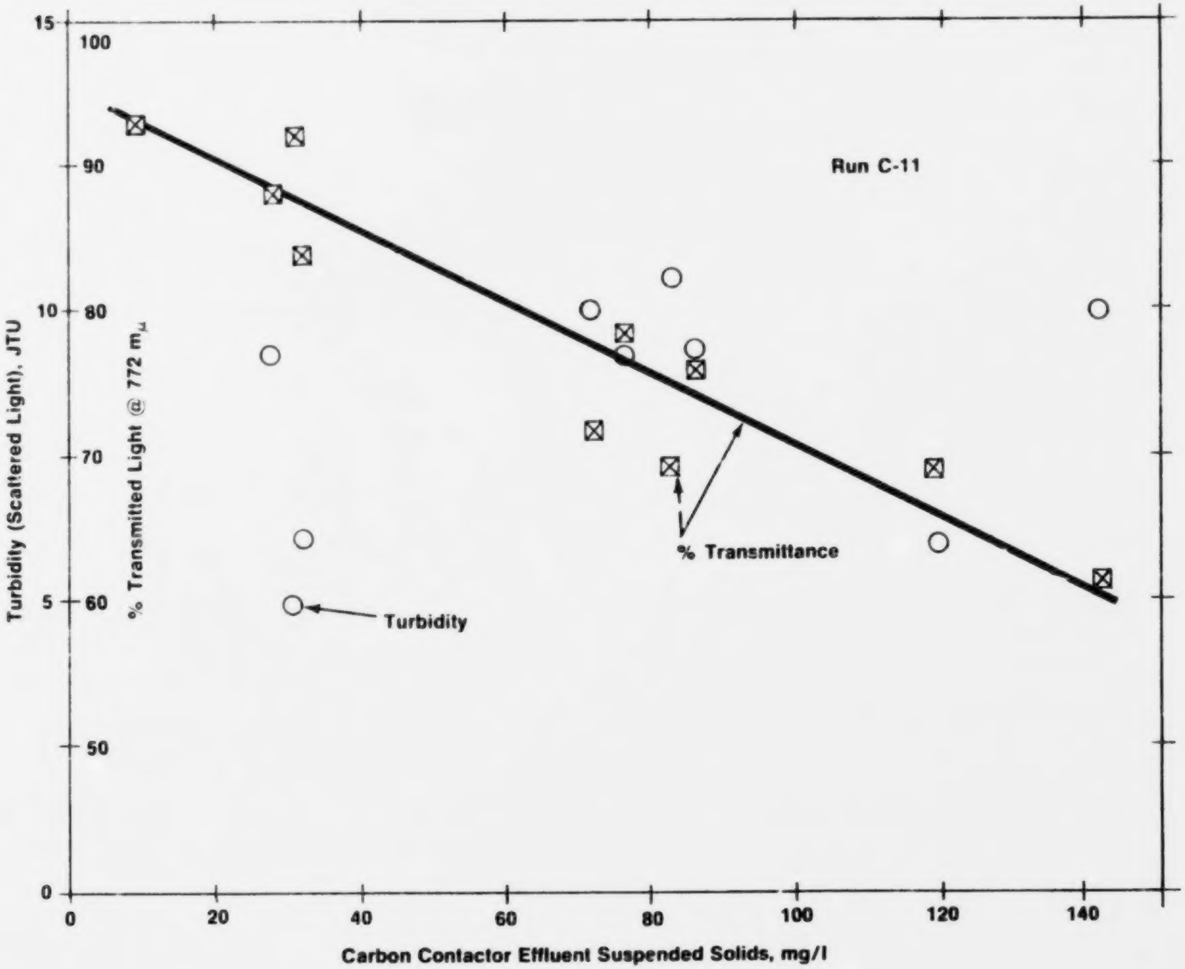
During pilot plant operation, evaluation of carbon contactor effluent clarity was made by three methods. First, the plant process turbidity system continuously monitored and recorded turbidity. The turbidity sensor used operated on a combination light scattering and transmitting principal. This device provided a useful instantaneous indication of clarity to operation personnel. Secondly, composite sample turbidity was determined with a 90° light scattering laboratory turbidimeter. A good correlation was found between 90° turbidity and suspended solids for virgin carbon suspensions in distilled water at concentrations below 30-40 mg/ℓ (see Figure B-1 in Appendix B). However, composite sample turbidity and suspended solids did not correlate, as typical data presented in Figure 39 show. The third method used, during the last few Runs (C-8 to C-12), was the measurement of transmittance of 772 mμ wave length light. Both laboratory calibration (see Figure B-1 in Appendix b) and plant data (see Figure 39) correlated well over the range of carbon contactor effluents encountered, It is, therefore, recommended that in future studies, carbon contactor effluent clarity be monitored by a light transmittance device.

TABLE 16: SULFIDE PROFILE THROUGH THE PAC-PCT PILOT PLANT

Run Number	Approximate SRT, Days		Sulfide Concentration,[a] mg/ℓ as $S^=$				
	First Stage	Second Stage	Raw Sewage	Chemical Effluent	First Stage Effluent	Second Stage Effluent	Plant (Filter) Effluent
C-2	2.4	2.2	0.76	0.35	0.46	0.96	0.36
C-3	2.7	4.5	0.10	0.02	0.02	<0.01	<0.01
C-5	4.6	--	0.11	<0.1	<0.1	--	<0.1
C-6	0.6	--	5.56	0.03	0.04	--	0.05
C-7	1.2	--	0.68	0.14	0.16	--	0.15

[a] Based on grab samples

FIGURE 39: CARBON TREATMENT:
INDIRECT EFFLUENT SUSPENDED SOLID MEASUREMENTS



Evaluation of carbon contactor effluent suspended solids data in Tables 13 and 14 indicates that they varied widely from Run to Run, with median values ranging from 3 to 53 mg/ℓ. Figures 40 and 41 show effluent suspended solids variation during two Runs.

It should be recalled that no arbitrary effluent quality parameters (e.g., suspended solids) were sought after during pilot plant operation. Rather operating conditions were specified and effluent quality determined for these conditions.

Carbon solids removal was accomplished by gravity sedimentation followed by granular media filtration. As will be shown in the next section of this report, the granular media filter effectively removed carbon solids, even at feed solids of 160 mg/ℓ. However, filter station efficiency, expressed as backwash water recycle requirements, suffered greatly at these high feed solids levels. Recognizing that granular media filtration efficiency is related to carbon contactor effluent suspended solids an attempt was made to quantify the effect of carbon contactor operating parameters on effluent solids.

Evaluation of hydraulic loading data indicated that the carbon contactors were operated at less than some critically high hydraulic loading which would have precluded good clarification. In general, the plant data was not precise enough to show a useful correlation between hydraulic loading and effluent suspended solids.

As discussed in Section IV, concentrated powdered carbon suspensions exhibit a "self flocculation" property. Since reaction-zone suspended solids concentration was representative of the total carbon contactor slurry concentration, and varied over a range of 2-21 g/ℓ, an attempt was made to correlate same with effluent solids A correlation was found for the single-stage carbon contacting Runs (Table 14), and is shown in Figure 42 (Reaction-Zone Solids for Runs C-11 and C-12 were estimated from 5-minute slurry settling test data). The data show that effluent suspended solids increased as reaction-zone solids concentration increased over the range of data shown. This observation does not necessarily refute the "self flocculation" property noted above in as much as some minimum concentration required may have been exceeded during all operating periods. The data in Figure

FIGURE 40: CARBON TREATMENT: EFFLUENT SUSPENDED SOLIDS

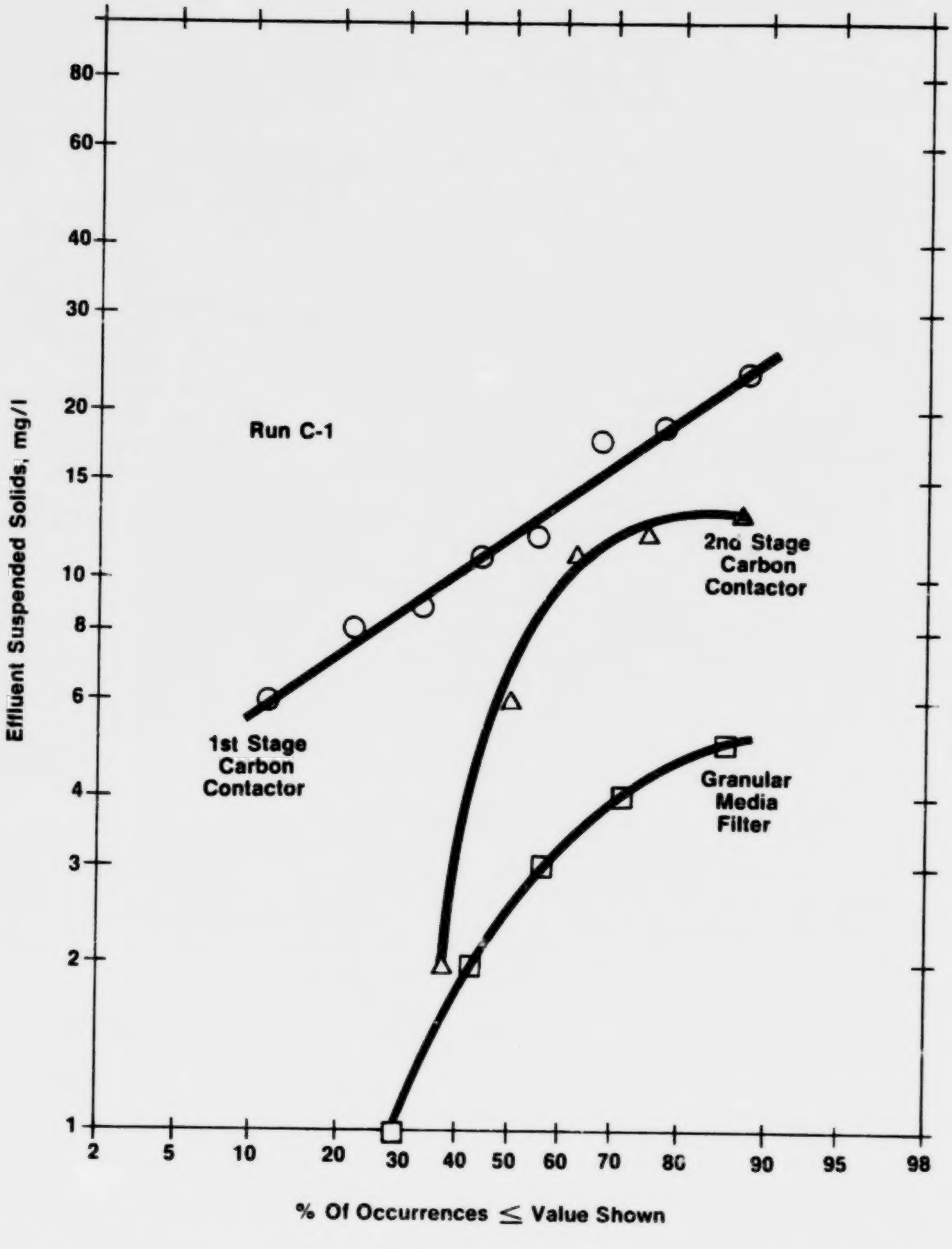


FIGURE 41: CARBON TREATMENT: EFFLUENT SUSPENDED SOLIDS

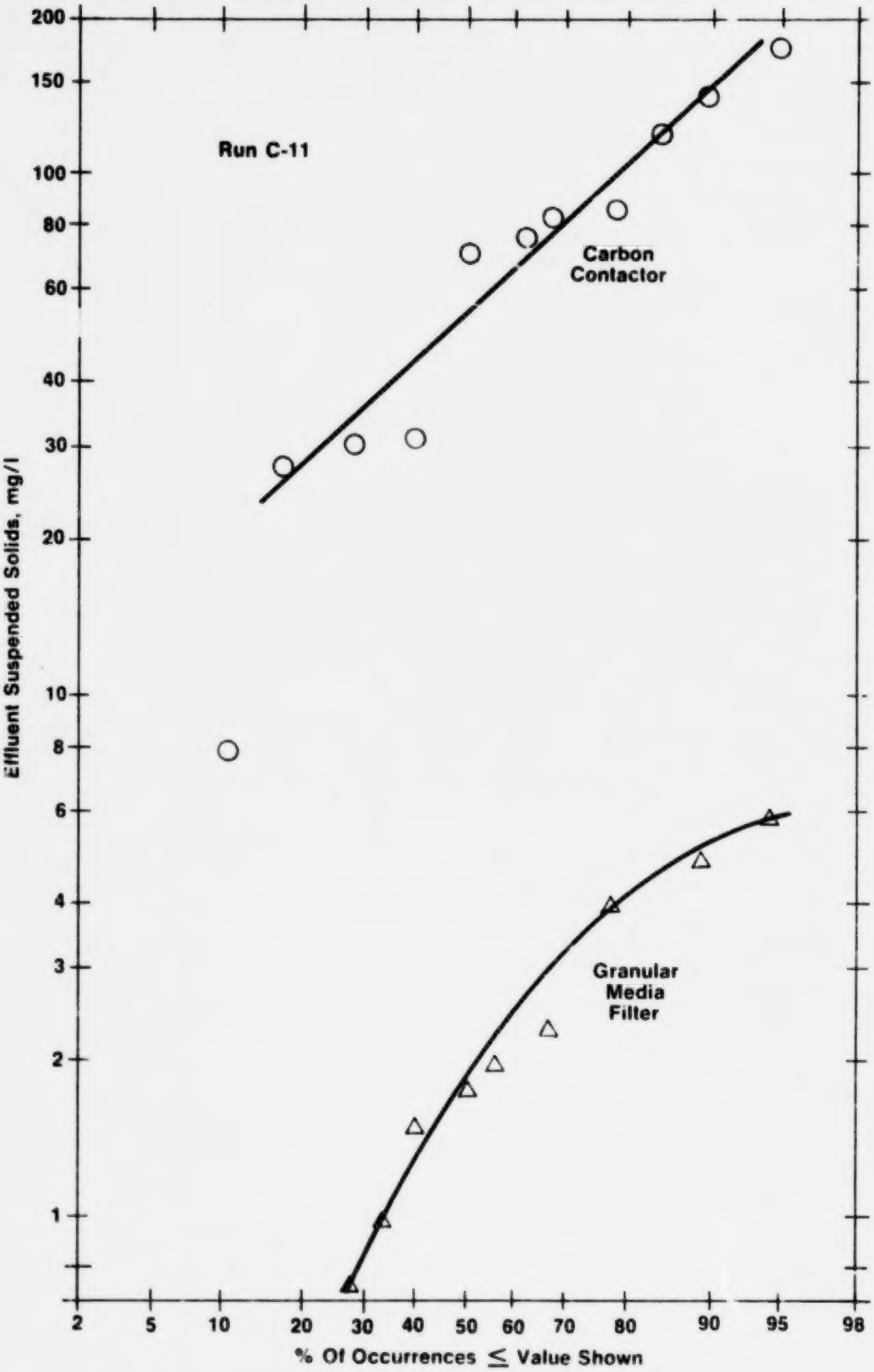
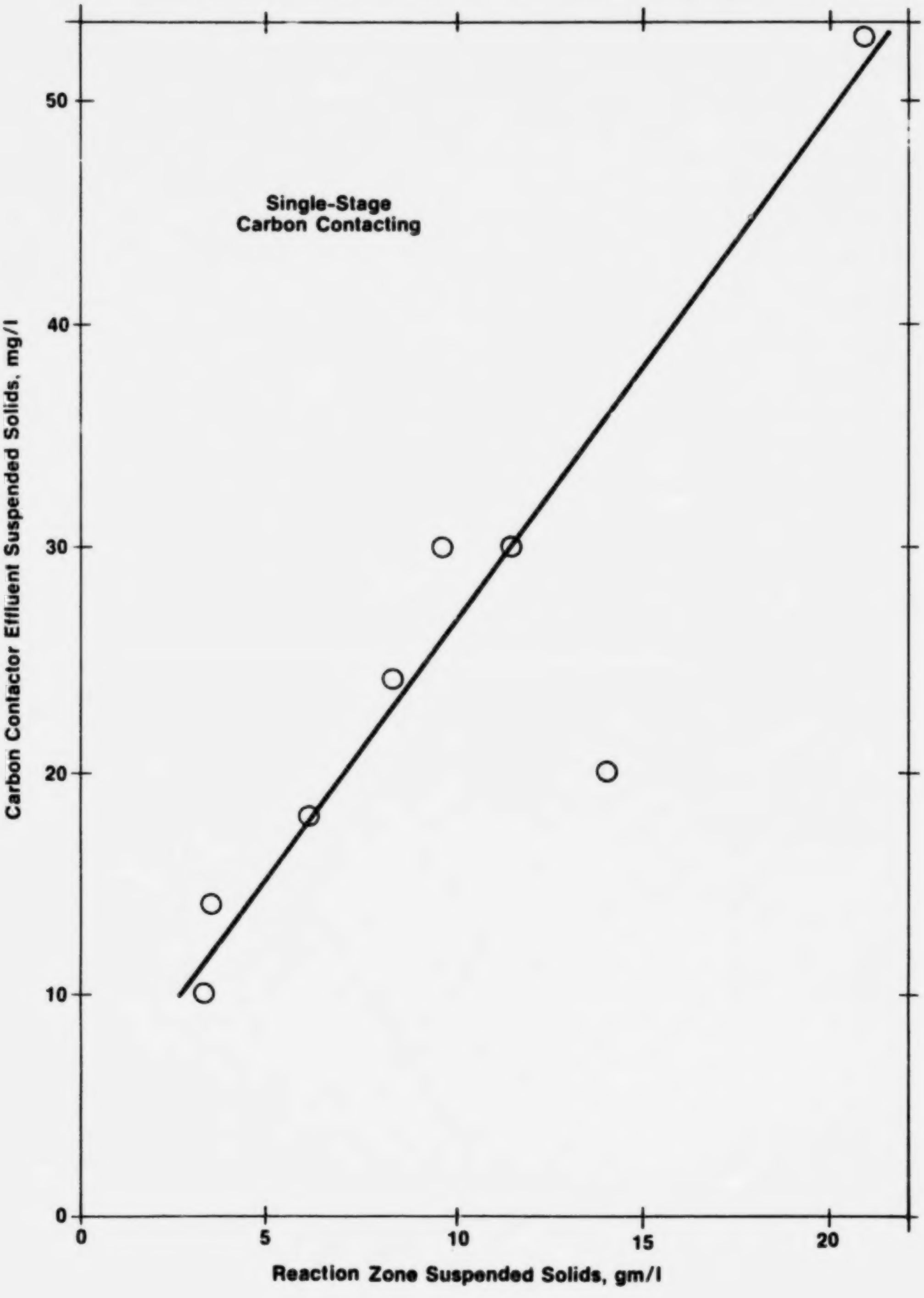


FIGURE 42: CARBON TREATMENT: EFFLUENT SUSPENDED SOLIDS



42 indicates that the solids-contact unit achieved a constant 99.7 percent removal of reaction zone suspended solids. Based on the relationship shown in Figure 42, one could conclude that operation at low reaction zone solids concentration is required if low effluent solids concentration is desirable. It will be shown later in this report that the levels of effluent solids shown in Figure 42 were readily removed by the granular media filter. Thus, a low concentration of carbon contactor effluent solids is not necessary to maintain low solids levels in the plant effluent.

Figure 43 is a typical plot of turbidity for all process streams through the pilot plant over a 24-hour period. The turbidity data is from the plant process turbidimeter print out. It is apparent that the chemical and carbon treatment clarifier effluent turbidities vary as their feed turbidities vary. Peak and minimum turbidities for the various streams are displaced by about the hydraulic retention time of the units. Positive liquid-solids separation was provided by the granular media filter resulting in production of a consistent plant effluent clarity. The similar trends in clarifier effluent turbidity prompted an analysis of the effect of feed solids on effluent solids. For two-stage counter-current contacting data in Table 13, there is a general trend of higher carbon contactor effluent suspended solids for higher feed suspended solids. However, a precise correlation does not exist. For the single-stage contacting data in Table 14, a fairly precise negative correlation was found between effluent and feed suspended solids. Figure 44 shows this correlation. It is significant to note that lower carbon contactor effluent solids values occurred with alum or iron chemical pretreatment (see Table 14). It is probable that the small amounts of alum and iron floc present in the carbon contactor feed aided flocculation of fine carbon particles.

A direct comparison of the effect of carbon contactor feed solids on effluent solids shown in Figures 43 and 44 indicates a contradiction. Though the data in Figure 43, is for only one days operation, it represents a generally observed system response. It is very probable that the response shown in Figure 44 is more a function of the type of suspended solids than the concentration of solids in the feed.

Others have reported the necessity of substantial amounts (1/2 - 1 mg/ℓ) of floccuation aids to effect good gravity

FIGURE 43: CARBON TREATMENT: TRANSIENT TURBIDITY EFFECT ON CLARIFICATION

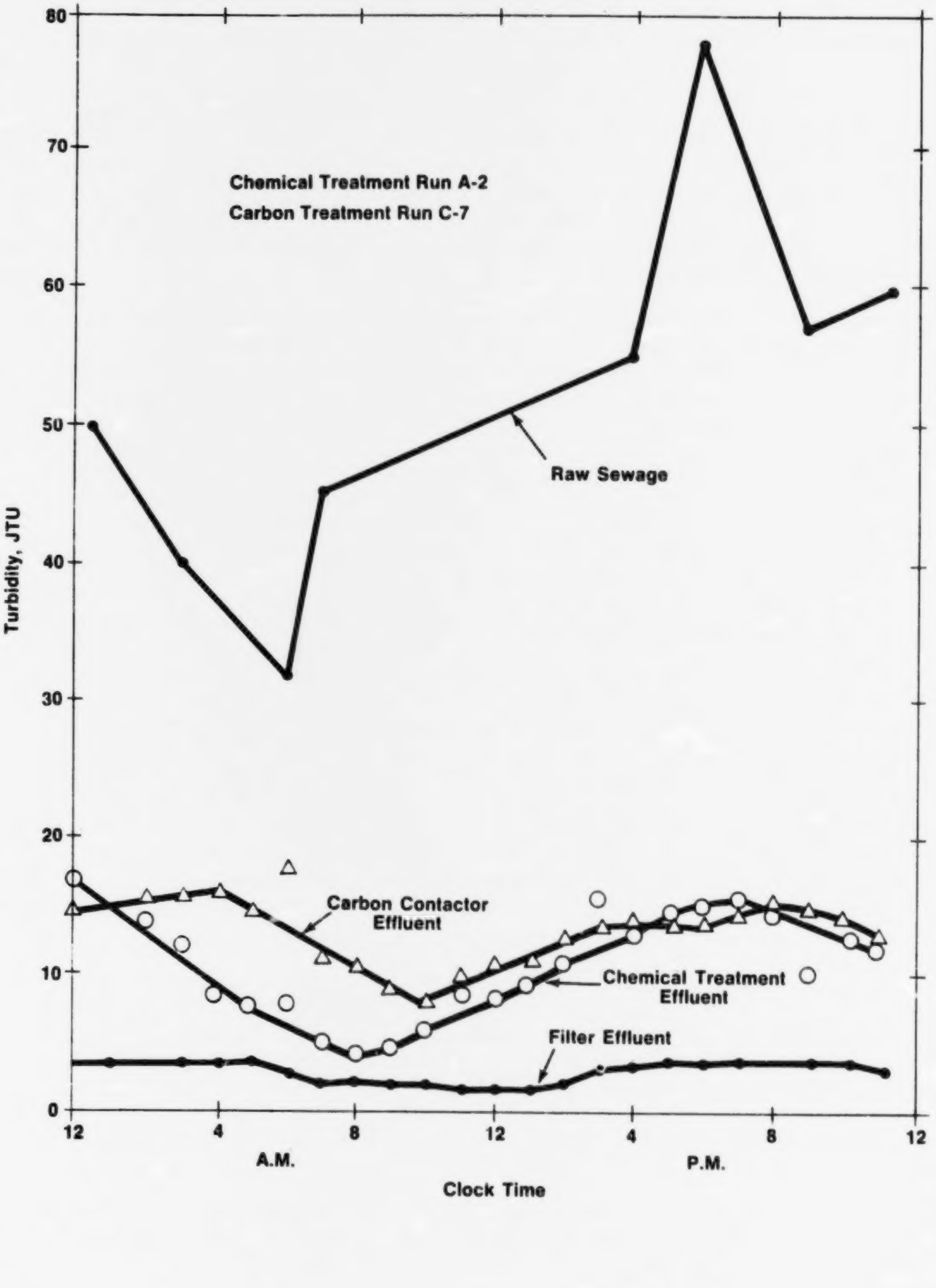
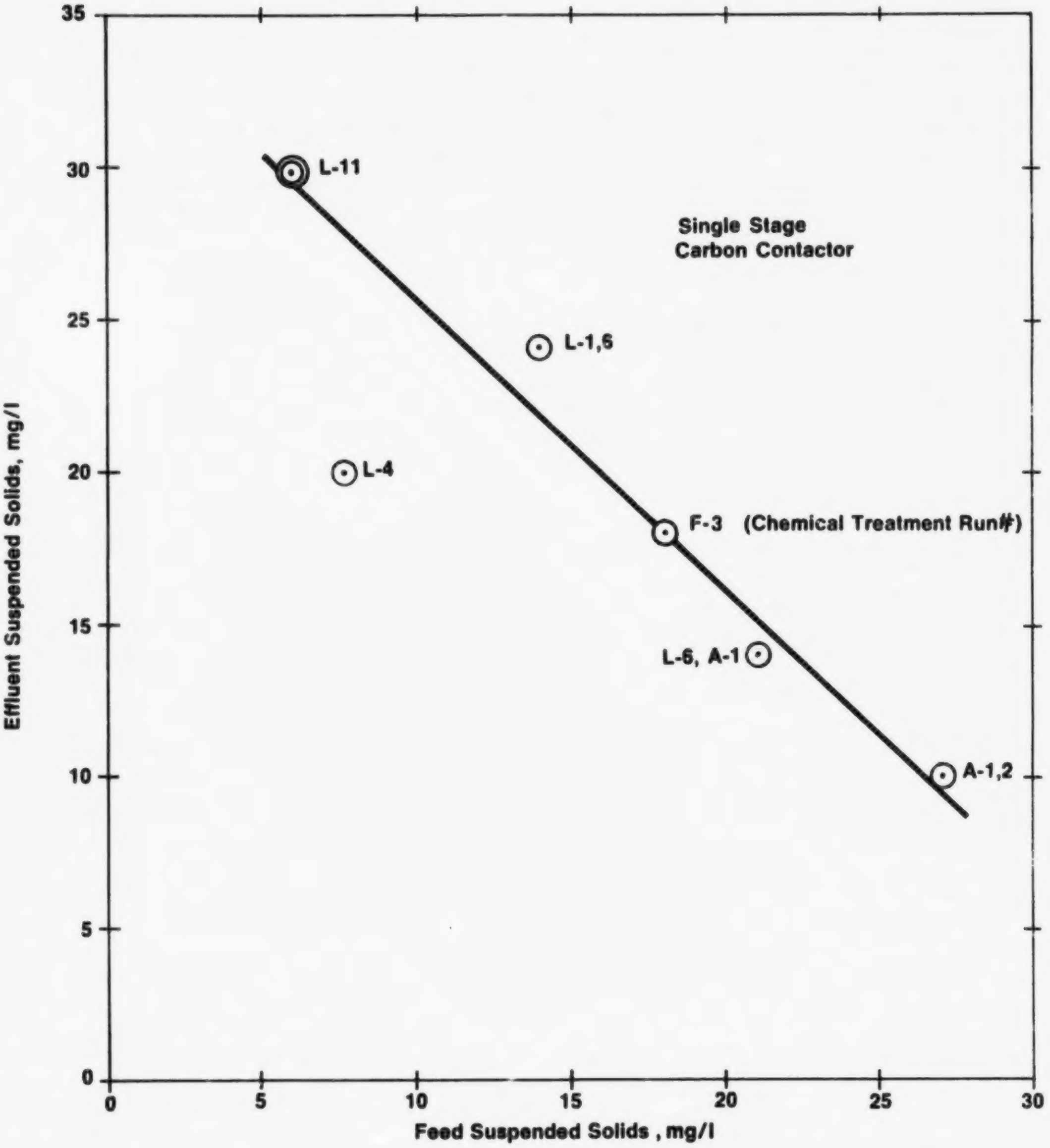


FIGURE 44: CARBON TREATMENT : FEED VERSUS EFFLUENT SUSPENDED SOLIDS



clarification of carbon suspensions[13,14,15,16]. During the first few months of plant operation, anionic polyelectrolyte was fed to the reaction zone of the carbon contactors at dosages ranging from 0.3 to 1.5 mg/ℓ. Clarification effectiveness was satisfactory as indicated by the data for Runs C-14 and C-2 in Table 13. During Run C-1, use of polyelectrolyte was discontinued and yet good clarification was experienced at a 0.76 gpm/sq ft hydraulic loading. An evaluation was made of the economic trade-off between polyelectrolyte cost and installed solids-contact treatment unit amortized cost. Table 17 shows this analysis. If acceptable clarification was obtained at a designed hydraulic loading of 0.48 gpm/sq ft without polyelectrolyte, then similar clarification would have to be achieved at a three-fold increase in hydraulic loading to justify use of only 0.14 mg/ℓ polyelectrolyte.

After Run C-2, no polyelectrolyte was used for aiding carbon flocculation. During Run C-11, excessive carbon carryover from the clarifier was experienced (as high as 180 mg/ℓ). Perhaps polyelectrolyte should have been used, however, the granular media filter effectively removed this excessive carbon carryover as the data in Figure 41 shows.

Polyelectrolyte feeding capabilities are recommended for the carbon clarification system used in this study, but polyelectrolyte should only be fed if plant effluent suspended solids are excessive or if inefficient filter operation is experienced. Filter operation is discussed in the next section.

In summary, it can be concluded that the 78 to 99 percent removal of powdered carbon plus feed suspended solids was an acceptable clarification effectiveness by the carbon contactors.

Granular Media Filtration

Mechanical operation of the 3.5 ft diameter plexiglass, automated filter was satisfactory. However, on two occasions, excessive pressure was applied to the filter housing resulting in failure of plexiglass-solvent bonds. Not withstanding the several weeks downtime and loss of data due to structural failures, the ability to visually observe filtering and backwashing action was very valuable. Visual observations were used to verify conclusions made from analysis of headloss data; for example, the existance of a schmutzdecke, or failure to remove same during backwashing, or penetration of carbon through the coal layer to the sand layer, or the existance of mudballs.

TABLE 17: COST TRADE-OFF OF POLYELECTROLYTE AND CLARIFICATION AREA FOR CARBON TREATMENT (10 MGD FLOW)

SOLIDS-CONTACT UNIT		POLYELECTROLYTE		
Overflow Rate gpm/sq ft	Amortized Capital Cost[a] (¢/1000 gal.)	Dose (mg/ℓ)	Chemical Cost[b] (¢/1000 gal.)	Total Cost (¢/1000 gal.)
0.48	0.39	0	0	0.39
0.97	0.33	0.08	0.06	0.39
1.46	0.28	0.14	0.11	0.39

a - 7 1/2% interest, 20 year for estimated erected equipment

b - @$1.50/lb of anionic polymer

The effectiveness of granular media filtration for clarifying carbon contactor effluent is evidenced by the consistantly low concentrations of suspended solids shown in Tables 13 and 14. With the exception of Run C-2 filter effluent suspended solids ranged from 1.3 to 3 mg/ℓ at hydraulic loadings of from 1.2 to 4.7 gpm/sq ft. The ability of the filter to remove high levels of feed suspended solids is indicated by data compiled during Run C-11 and shown in Table 18. It is apparent, however, that very short filter cycles were experienced for the high carbon concentration applied, resulting in high volumes of backwash recycle.

Figure 45 shows the effect of filtration rate on cycle time when low filter feed suspended solids were encountered. The dashed curve was constructed to show the filtration rate - cycle time relationship that results in a five percent backwash recycle. The region above and to the right of the dashed curve represents less than 5 percent backwash recycle. Water backwashing consisted of 26 gpm/sq ft for 6 minutes after air-scour at 4.5 to 5.5 SCFM/sq ft for 1.5 minutes (see Appendix D).

The backwash rate, or upflow velocity required to expand the filter bed and allow flushing out of solids is governed by the media size and density, and the water temperature. An acceptable bed expansion of about 25 percent was attained at 26 gpm/sq ft. The duration of backwash is governed by the geometry of the filter housing. The pilot plant filter had a freeboard of 1 1/2 times the bed depth. Most commercial designs have a freeboard of only 1/2 to 3/4 times the filter bed depth. Thus, the volume of water to be displaced from the pilot plant filter chamber was 65 to 110 percent more than for most commercial designs. The backwash conditions used during this study (26 gpm/sq ft @ 6 minutes) amounted to displacement of five (5) volumes of filter chamber liquid. Several commercial filters are designed to accomplish satisfactory flushing of dislodged solids with displacement of approximately three (3) volumes of filter chamber liquid. Needless to say, extrapolation of backwash water recycle requirements to full scale operation must take into consideration the characteristics of the specific filtration equipment considered. Based on the above factors of filter chamber volume and number of displacements, it is probable that about 30 to 40 percent less backwash recycle than experienced in this study would be required for full scale filtration systems.

TABLE 18: GRANULAR MEDIA FILTRATION: CLARIFICATION EFFECTIVENESS AT HIGH FEED SOLIDS

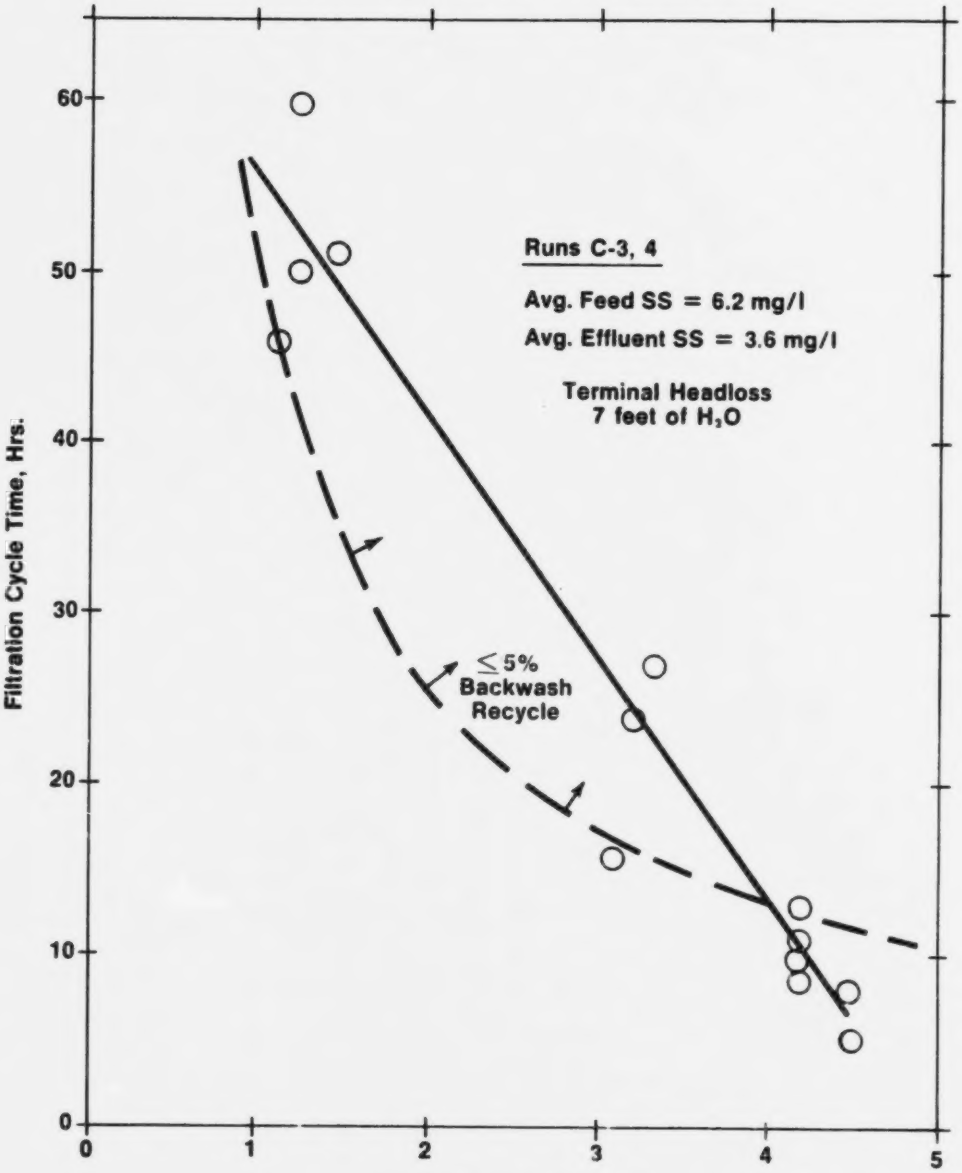
(Averages of data obtained during Run C-11)

# of Filter Cycles	Suspended Solids, mg/ℓ		Filtration Rate gpm/sq ft	Cycle Time[a] hr	Backwash Recycle[b] %
	Feed	Effluent			
5	31	0.8	5.0	5.7	9
14	76	5.0	1.9	7.0	20
7	160	2.0	2.1	5.3	23

a - To a terminal headloss of 7 ft of water

b - H_2O backwash @ 26 gpm/sq ft for six minutes

FIGURE 45: GRANULAR MEDIA FILTRATION: CYCLE TIME vs. FILTRATION RATE



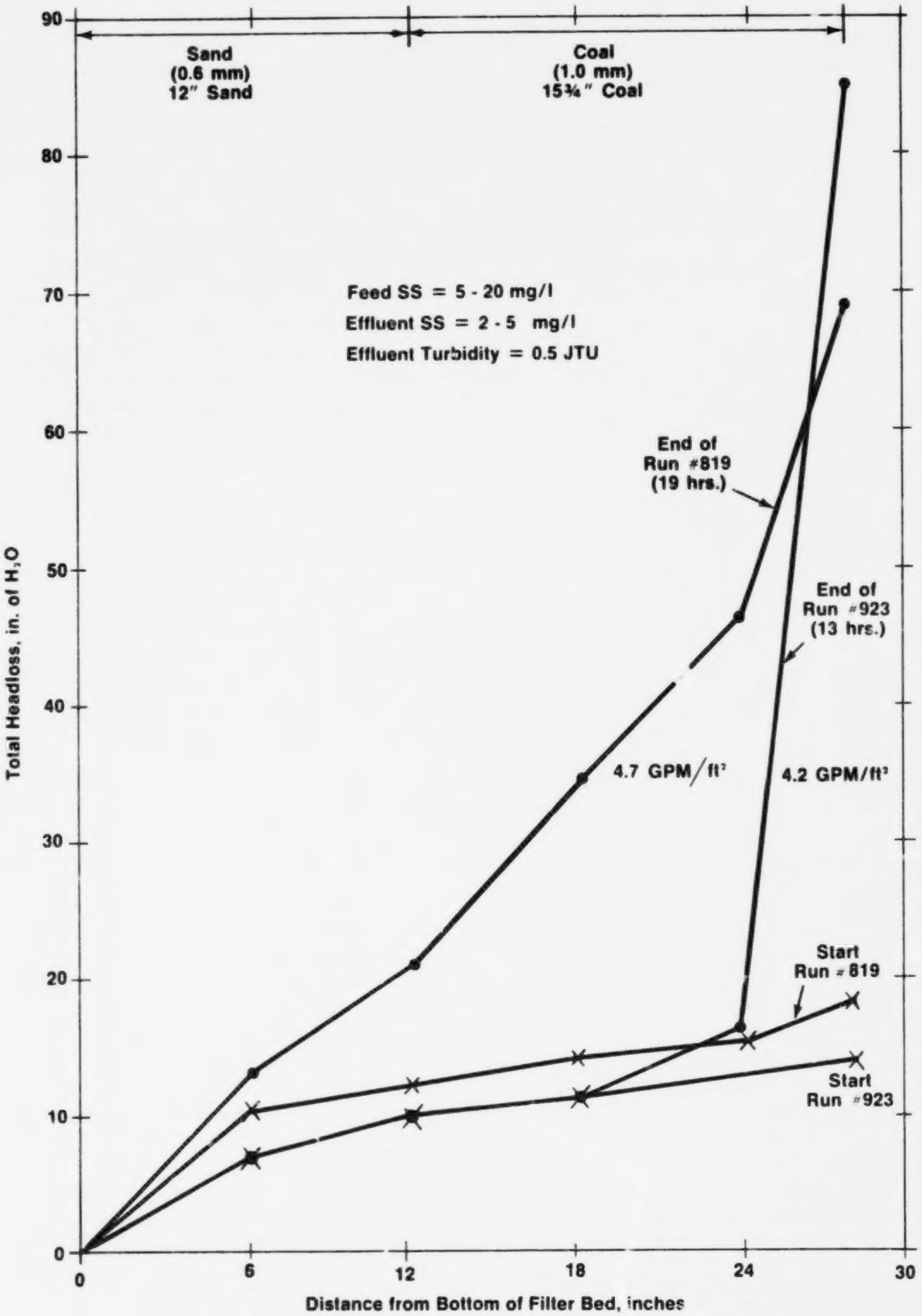
Evaluation of incremental headloss data measured at six (6) inch height increments throughout the filter bed depth indicated a deficiency in the filter bed design used. Figure 46 shows headloss profiles at the start and end of two filter cycles during Runs C-1 and C-2. The headloss profile at the end of filter Run 923 is representative of about 90 percent of all filter Runs analyzed. Essentially all the headloss occurred over the top few inches of the coal layer indicating that the size of coal (1.0 mm) was too small to allow penetration of carbon particles, even at filtration rates of 4-5 gpm/sq ft. The headloss distribution at the end of Run 819 (Figure 46) was an exception and shows a much more desirable headloss pattern. The headloss is uniformly distributed throughout most of the filter bed depth. No apparent reason was found for the desirable headloss pattern found in Figure Run 819. The significance of obtaining a more desirable headloss distribution is indicated by comparing the volume of water filtered and backwash water recycle requirements. Some 64 percent more water was filtered during Run 819 compared to Run 923 and, hence, 40 percent less backwash recycle water was required.

Because the carbon particles were removed at the top surface of the coal layer, use of 0.1 to 0.3 mg/ℓ of anionic polyelectrolyte floc strengthener during a few filter Runs produced no discernable benefit of improved headloss distribution or filtrate clarity. For the same reason, filtration of feed suspended solids as high as 160 mg/ℓ resulted in no deterioration in filtrate clarity. Obviously, once a schmetzdecke of carbon particles is formed, the existance of "tough" floc or high feed solids concentrations will not affect effluent clarity or headloss distribution.

Six (6) times during the course of this study, about 1 to 1 1/2 inches of coal media was removed from the filter bed surface and replaced with new coal media. It was hoped that removal of coal fines would allow penetration of carbon particles. No significant improvement in headloss distribution was observed.

During Runs C-3 and C-4, the filter was operated for several filter cycles at a filtration rate of about 1.3 then 3.2 and finally at 4.2 gpm/sq ft. At 1.3 gpm/sq ft, cycle times were about 50 hours as seen in Figure 45. Under these operating conditions, numerous "mud balls" were observed, ranging in size from 1/2 to as large as 1 1/2 inches in

FIGURE 46: GRANULAR MEDIA FILTRATION: HEADLOSS DISTRIBUTION



major dimension. The normal air-scour, water backwashing procedure did not eliminate the "mud balls". However, when the filtration rate was increased to 3.2 gpm/sq ft, the cycle time was reduced to about 22 hours and the mud balls were completely eliminated by the normal backwash procedure. Evidently, detention of carbon material on the filter bed surface for 50 hours results in "cementation" of the solids whereas detention for 22 hours did not. It is possible that biological slimes were produced at 50 hours carbon detention time, but not at 22 hours carbon detention time. Regardless of the cause, it is suggested that the granular media filter be programmed to backwash at least once per day to minimize the problem of mud balls.

The results presented and discussed show that carbon contactor effluent was effectively clarified by granular media filtration. Additional studies are needed to develop improved efficiency of filter operation. It is recommended that coarser coal media be evaluated to determine a size which will allow significant penetration of carbon particles to the underlying sand layer. The effect of polyelectrolyte can then be determined and optimum filtration rates established. It is also recommended that future studies be designed to verify or define near optimum backwashing procedures.

Removal of Soluble Organics

The singular purpose of powdered carbon treatment was removal of soluble organics from the wastewater. Thus, SCOD, STOC and carbon dosage data from Tables 13 and 14 were analyzed in an attempt to determine the relationship between carbon system feed and effluent soluble organics and carbon dosage. For obvious reasons, the applicability of an adsorption model was evaluated.

Tables 19 and 20 present observed median SCOD and STOC data and reduced data which indicate incremental organic reductions, organic removals and feed organic loadings. During pilot plant operation, granular media filtration of all carbon contactor effluent was impossible (e.g., when carbon contactors were run in parallel). Thus, the effluent SCOD and STOC concentration shown are for the carbon contactors and not the granular media filter.

All STOC and SCOD data were analyzed to ascertain whether or not a consistant and precise correlation existed between the two. Table 21 shows the results of linear

TABLE 19: CARBON TREATMENT: SUMMARY OF SCOD REMOVAL DATA
(Two-Stage Counter-Current)

OBSERVED DATA					REDUCED DATA							
		SCOD, mg/ℓ			SCOD Removed, mg/ℓ			Organic Removal g/g			Organic Loading $\left(\frac{C_o}{M}\right)$, mg/ℓ/mg/ℓ	
Run #	Carbon Dosage (M) mg/ℓ	Feed (C_o)	Inter-mediate Stage (C_i)	Effl. (C_e)	1st Stage (X")	2nd Stage (X')	All Stages (X)	All Stages $\frac{X}{M}$	2nd Stage $\frac{X'}{M}$	1st Stage $\frac{X''}{M}$	1st and All Stages	2nd Stage
C-1	325	50	12	12	38	0	38	0.12	0.000	0.12	0.15	0.04
C-2	150	43	21.5	18	21.5	3.5	25	0.17	0.023	0.14	0.29	0.14
C-8	150	45	18	11	27	7	34	0.23	0.047	0.18	0.30	0.12
C-3	110	68	28	22	40	6	46	0.42	0.055	0.36	0.63	0.26
C-4	75	70	38	30	3.2	8	40	0.53	0.107	0.43	0.91	0.50

TABLE 19 (cont.): CARBON TREATMENT: SUMMARY OF SCOD REMOVAL DATA
(Single-Stage)

OBSERVED DATA				REDUCED DATA		
		SCOD, mg/ℓ		SCOD Removed mg/ℓ	Organic Removal g/g	Organic Loading $\left(\frac{C_o}{M}\right)$ mg/ℓ/mg/ℓ
Run #	Carbon Dosage (M) mg/ℓ	Feed (C_o)	Effl. (C_e)	All Stages (X)	All Stages $\frac{X}{M}$	All Stages
C-11	345	41	6.4	34.6	0.099	0.12
C-12	317	41	13	28	0.086	0.13
C- 6	300	40	12	28	0.093	0.13
C- 9	146	35	25	12	0.076	0.24
C-10	146	35	23.5	10	0.068	0.23
C- 7	105	30	17	13	0.124	0.29
C- 5	95	58	22	36	0.380	0.61

TABLE 20: CARBON TREATMENT: SUMMARY OF STOC REMOVAL DATA

OBSERVED DATA					REDUCED DATA					
		STOC, mg/ℓ			STOC Removed, mg/ℓ		Organic Removal g/g		Organic Loading $\left(\frac{C_o}{M}\right)$, mg/ℓ/mg/ℓ	
Run #	Carbon Dosage (M) mg/ℓ	Feed (C_o)	Inter-mediate Stage (C_i)	Effl. (C_e)	2nd Stage (X')	All Stages (X)	All Stages $\frac{X}{M}$	1st Stage $\frac{X'}{M}$	1st And All Stages	2nd Stage
					Two-Stage Counter-Current					
C-2	150	22	15	12	3	10	0.067	0.020	0.15	0.10
C-8	150	24	11	9	2	15	0.10	0.013	0.16	0.07
					Single-Stage					
C-6	300	25	--	13	--	12	0.040	--	0.083	--
C-7	150	16	--	13	--	3	0.029	--	0.15	--
C-5	95	28	--	10	--	18	0.19	--	0.29	--

TABLE 21: CORRELATION OF SOLUBLE TOC WITH SOLUBLE COD

Carbon Treatment Run #	Effluent From	Linear Regression Equation	Correlation Coefficient
C-2	Chemical Contactor	STOC = 13 + 0.19 SCOD	0.61
	First Stage Carbon Contactor	STOC = 12 + 0.16 SCOD	0.47
	Second Stage Carbon Contactor	STOC = 11 + 0.10 SCOD	0.31
	GM Filter	STOC = 12 - 0.018 SCOD	0.09
C-7	Chemical Contactor	STOC = 2.7 + 0.49 SCOD	0.98
	Carbon Contactor	STOC = 2.0 + 0.56 SCOD	0.98

regression analysis of STOC and SCOD. Data from Run C-7 are precisely correlated and indicate that from 2.0 to 2.7 mg/ℓ STOC would exist if the SCOD were zero. The slope, or ratio of STOC/SCOD of about 0.5 is within the range of data by others[29,30]. The data from Run C-2 did not correlate precisely, but do seem to indicate that about 12 mg/ℓ of STOC would exist if the SCOD were zero. Data from Run C-6 correlates similarly to Run C-2, whereas data from Runs C-5 and C-7 did not correlate. It has been reported that silicates will interfere with the TOC analysis[31]. Silicate concentrations of 10 to 50 mg/ℓ (as SiO_2) were reported to register as 3 to 4 mg/ℓ TOC. Analysis of Salt Lake City tap water showed the presence of 10-15 mg/ℓ of SiO_2. Thus, STOC data presented in this report are probably high by about 3 mg/ℓ of STOC.

Application of Adsorption Model

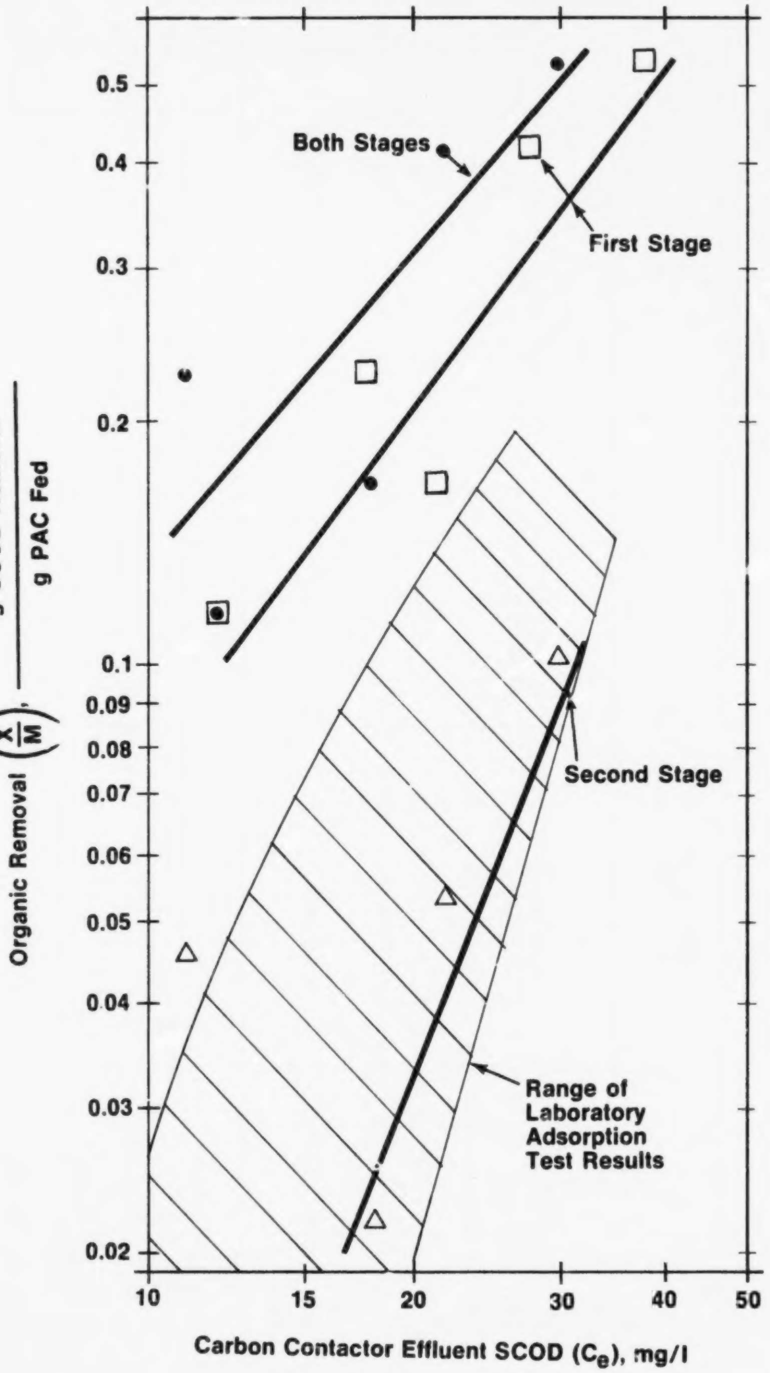
Correlation of SCOD removal data with the Freundlich math model (see Appendix A) was not very precise as indicated by regression analysis results in Table 22. As shown, however, a more precise correlation exists when organic removal (X/M, g of SCOD removed per g of carbon fed) was regressed on effluent SCOD (C_e) rather than the log of organic removal on the log of effluent SCOD as for the Freundlich math model.

Figure 47 is a plot of organic removal data for two-stage counter-current contact Runs, with regression curves. Approximately, 80 to 88 percent of the SCOD was removed in the first-stage, with the exception of Run C-1, where all measured SCOD removal occurred in the first-stage. This level of removal in the first-stage of contacting is considerably higher than would be predicted by the Freundlich model. As discussed in Appendix A, adsorption theory indicates that the first-stage and second-stage curves should be identical if the physical and chemical properties of adsorbed organics in both stages are identical. Considering the heterogeneous nature of soluble wastewater organics, it is probable that certain species are preferentially adsorbed in the first-stage of contacting. Thus,

TABLE 22: CARBON TREATMENT: ADSORPTION MODELS OF SCOD REMOVAL DATA

Contact Stage	Log $\frac{X}{M}$ vs Log C_e (Freundlich Model)		$\frac{X}{M}$ vs C_e	
	Regression Equation	Correlation Coefficient	Linear Regression Equation	Correlation Coefficient
Single-Stage	$\frac{X}{M} = 0.013{C_e}^{0.90}$	0.68	$\frac{X}{M} = -0.085+0.017C_e$	0.80
Two-Stage Counter-Current (2SCC)	$\frac{X}{M} = 0.0087{C_e}^{1.2}$	0.80	$\frac{X}{M} = -0.079+0.020C_e$	0.89
First Stage of 2SCC	$\frac{X}{M} = 0.0043{C_i}^{1.3}$	0.93	$\frac{X}{M} = -0.10+0.017C_i$	0.95
Second Stage of 2SCC	$\frac{X'}{M} = 3.3x10^{-6}{C_e}^{3.1}$	0.71	$\frac{X'}{M} = -0.033+0.0043C_e$	0.79
Single-Stage plus 2SCC	$\frac{X}{M} = 0.0087{C_e}^{1.1}$	0.71	$\frac{X}{M} = -0.11+0.020C_e$	0.87

FIGURE 47: CARBON TREATMENT: SCOD REMOVAL FOR TWO-STAGE COUNTER-CURRENT TREATMENT



it would be unrealistic to expect the observed first and second-stage results to be described by a precisely similar model. It would be just as unrealistic, however, to assume that the dramatic difference between the organic removal for the two stages could be explained away as being due to "preferential adsorption" in the first stage. A much more plausible explanation for the disporportionally high organic removal in the first-stage would be that biological oxidation of soluble organics was occurring. Qualitative indications of the presence of anaerobic biological activity were alluded to earlier in this section.

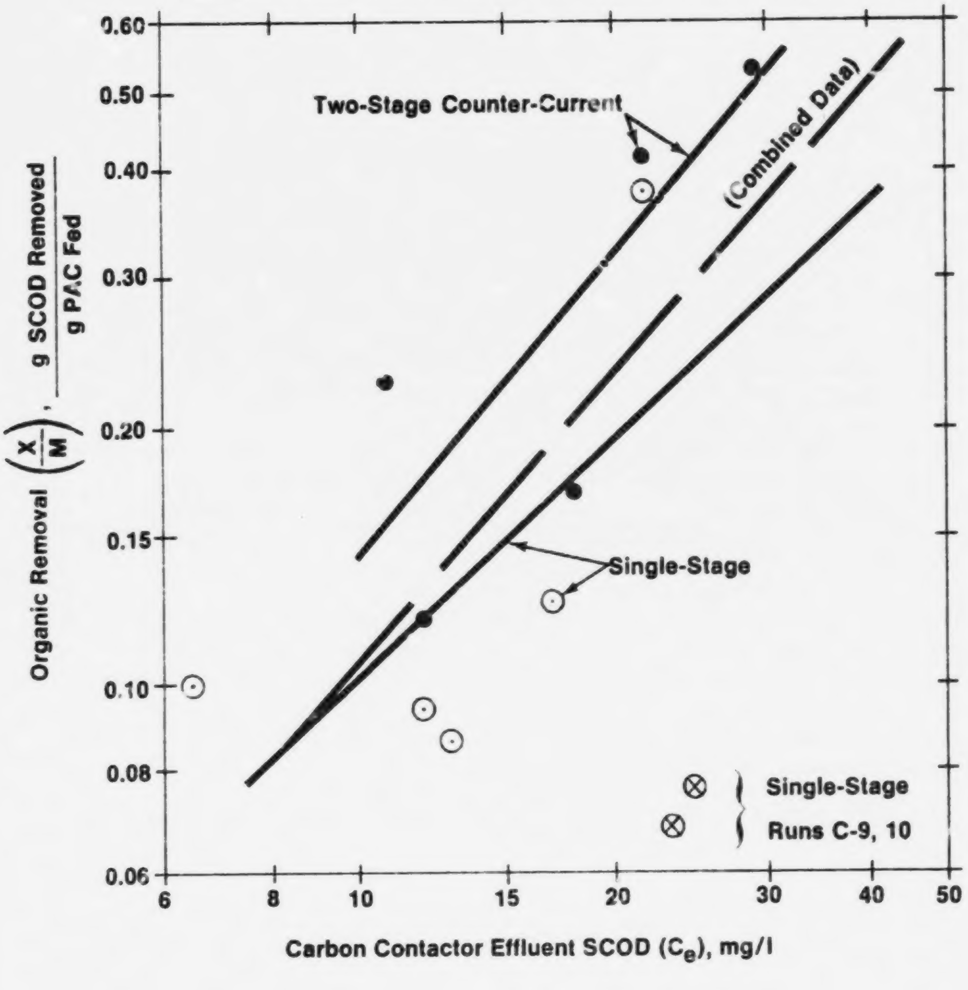
The cross-hatched area in Figure 47 represents the range of laboratory adsorption equilibrium isotherm test results. Organic removal during these tests was by adsorption only. The similar results for organic removal in the second-stage contactor and for the laboratory adsorption tests would seem to indicate that the major removal mechanism in the former was also adsorption. Organic removal in the first-stage contactor is seen to be considerably higher than predicted for adsorption by the laboratory adsorption test results. This fact reinforces the thesis that an organic removal mechanism in addition to adsorption existed in the first-stage carbon contactor.

Figure 48 shows SCOD removal results for both single-stage and two-stage counter-current contacting. It appears that organic removals, X/M, for two-stage treatment are generally higher than for single-stage treatment.

Evaluation of carbon SRT and type or level of chemical pretreatment conditions failed to reveal any consistant or definable effects on organic removal or effluent SCOD concentrations. Either no effects existed or the data, as modeled by the Freundlich equation, is not precise enough to define the effects.

The SCOD removal data from Runs C-9 and C-10 were not included in the above comparison of single-stage and two-stage treatment results. These two Runs were specifically conducted in an effort to quantify the effect of biological activity on SCOD removal. The two carbon contactors were operated in parallel at similar carbon dosages, overflow rates and carbon SRT's (see Table 14).

FIGURE 48: CARBON TREATMENT: SCOD REMOVAL FOR SINGLE AND TWO-STAGE COUNTER-CURRENT TREATMENT



A copper sulfate solution was fed to one contactor (Run C-10) at a dosage of 2.4 mg/ℓ of Cu^{++}. A laboratory equilibrium adsorption test was conducted to determine whether or not the addition of 20 mg/ℓ of Cu^{++} would affect the adsorption of SCOD from neutralized chemical contact effluent. The presence of Cu^{++} did not affect organic removal, X/M, in the equilibrium SCOD range tested (15-50 mg/ℓ). The presence of 2.4 mg/ℓ Cu^{++} in the feed to the carbon contactor was considered to eliminate biological activity.

During Run C-9 (concurrent with C-10), no Cu^{++} was fed, and a relatively long carbon SRT of 5.9 days was established in an effort to promote biological activity. However, no quanitative or qualitative indications of the presence of biological activity were experienced. Inspection of plant records indicated that during Runs C-9 and C-10 (chemical pretreatment Run L-11) the carbon contacting system feed pH was extremely erratic. The pH varied from 2 to 11 at frequencies of 1/2 to 3 hours. For two days the pH was about 10.7 due to a breakdown of the acid feeding system. Because of the widely varying pH during Run C-9, it is extremely doubtful that any significant level of biological activity was ever established. Thus, organic removal results from both Runs C-9 and C-10 were considered to be representative of non-biologically enhanced systems. The significance of allowing natural anaerobic process to occur is indicated by the fact that organic removals experienced in Runs C-9 and C-10 are only 1/3 of that predicted by the single-stage regression curve in Figure 48. Unfortunately, this observation is not absolutely conclusive in as much as the widely varying pH experienced during Runs C-9 and C-10 may have affected organic removal by adsorption. The effect of pH on organic removals for PCT carbon adsorption systems has not been established, or reported in the literature.

The existance of a biological organic removal mechanism is indicated by data from Run C-13. During this Run no virgin carbon was fed to the contactor. Effluent carbon was captured by the granular media filter and recycled to the carbon contactor. During this 13 day Run about 30 percent removal of COD and TOC was observed. At an overflow rate of 0.75 gpm/sq ft, the wastewater - powdered carbon slurry contact time was less than 30 minutes, while the carbon SRT was indefinitely long.

The STOC organic removal data in Table 20 are contrary to what adsorption theory would predict. For both single-stage and two-stage contacting, higher organic removals, X/M, are associated with lower equilibrium STOC concentrations, C_e.

Alternate Organic Removal Model

Recognizing the inappropriateness of describing the pilot plant results by an "adsorption" model several attempts were made to more precisely relate organic removal and carbon dosage. The following emperical relationship, which was suggested by the project officer, was found to precisely describe the results:

$$X/M = K_1 + K_2 \frac{C_o}{M} \qquad (6)$$

Where: X/M = mg/ℓ SCOD removed in any given stage per mg/ℓ of carbon.

$\frac{C_o}{M}$ = mg/ℓ feed SCOD to any given stage per mg/ℓ of carbon.

K_1, K_2 = constants

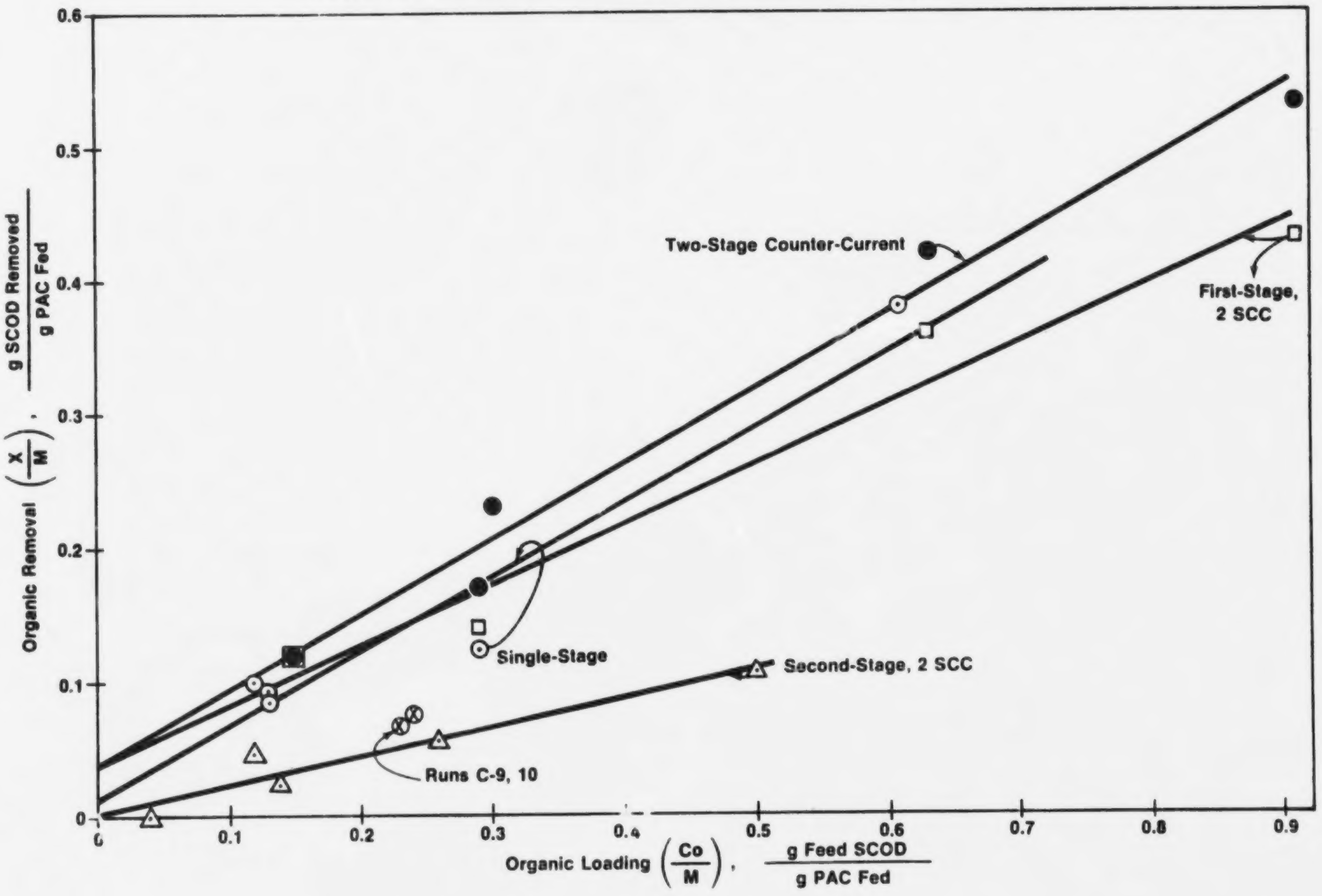
Table 23 presents linear regression equations and correlation coefficients for SCOD removal results. Correlation coefficients of 0.97 to 0.99 indicate a presice fit of the data. A linear plot of X/M vs C_o/M for all stages of treatment is shown in Figure 49.

A statistical analysis of significance using the student's "t" distribution (see Reference 32) resulted in accepting the hypothesis that the slopes, K_2, of the single-stage, two-stage counter-current and combined data were not significantly different. Student "t" values of 0.4 and 0.9 were computed for 3 and 8 degrees of freedom respectively. To determine if the height of the single-stage curve in Figure 49 was significantly different than for the two-stage counter-current curve, an estimate of X/M

TABLE 23: CARBON TREATMENT: MODEL OF SCOD REMOVAL DATA

Contact Stage	Linear Regression Equation	Correlation Coefficient
Single-Stage	$\frac{X}{M} = 0.0116 + 0.557 \frac{C_o}{M}$	0.97
Two-Stage Counter-Current (2SCC)	$\frac{X}{M} = 0.0397 + 0.558 \frac{C_o}{M}$	0.99
First Stage of 2SCC	$\frac{X''}{M} = 0.0413 + 0.445 \frac{C_o}{M}$	0.99
Second Stage of 2SCC	$\frac{X'}{M} = 0.00070 + 0.216 \frac{C_i}{M}$	0.97
Single-Stage plus 2SCC	$\frac{X}{M} = 0.0179 + 0.579 \frac{C_o}{M}$	0.99

FIGURE 49: CARBON TREATMENT: ORGANIC REMOVAL MODEL (SCOD)



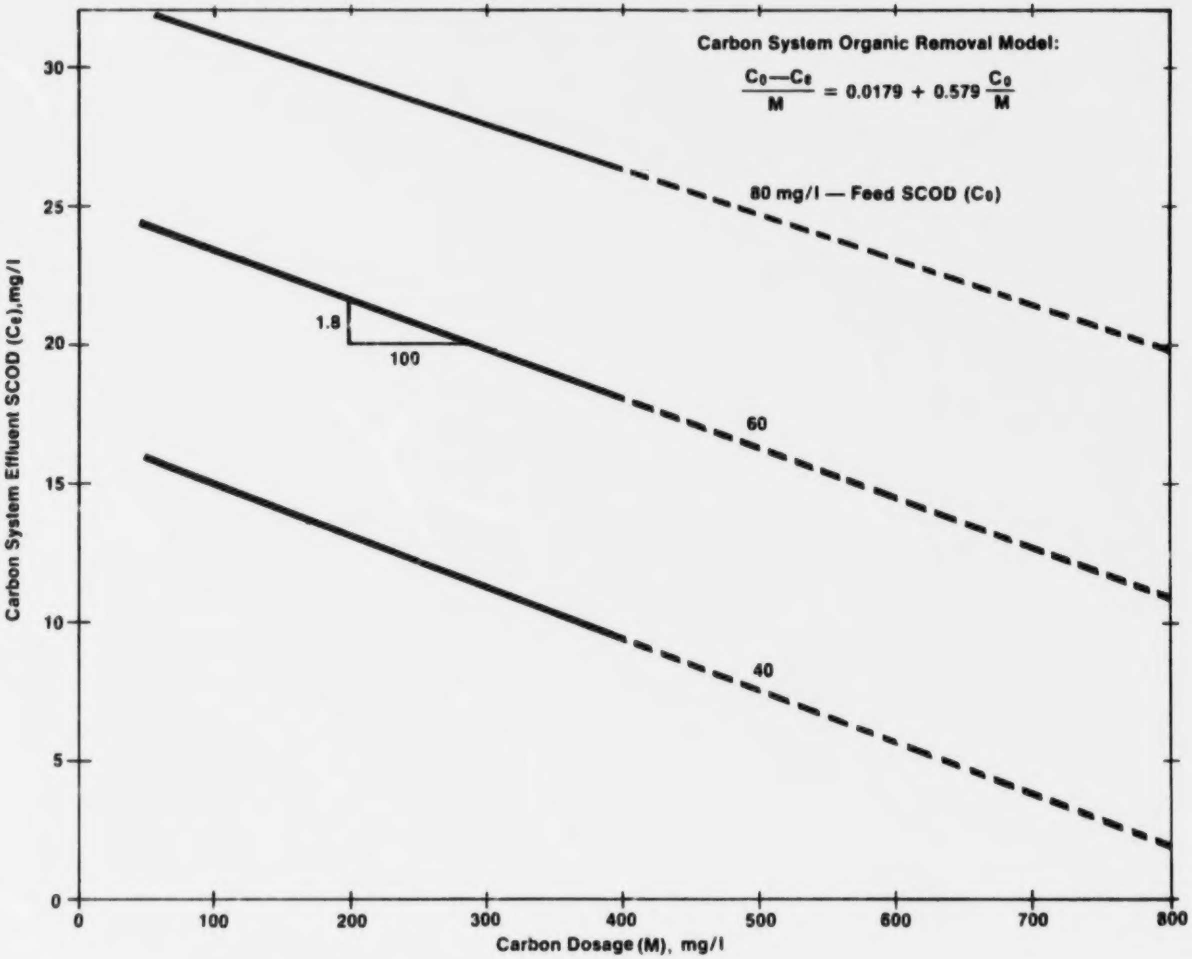
at the combined average C_o/M of 0.36 was determined from the regression equations. This average value of C_o/M was chosen because the variance of the estimated X/M's would be a minimum at this point. The hypothesis that the estimated X/M's (0.21 and 0.24) were not significantly different resulted in a student's "t" of about 1.7 at 3 degrees of freedom. The hypothesis was marginally accepted. Had the variance of the estimated X/M's been slightly less, the hypothesis would have been rejected and the height of the regression curves considered significantly different. Combination of single-stage and two-stage counter-current data reduced the variance of estimated X/M's by approximately 60 percent.

Based on the above statistical analysis, it was concluded that the variability of pilot plant results precluded quantifying the difference between X/M values for single-stage and two-stage counter-current contacting, with a significant degree of statistical confidence. Additional pilot plant studies should provide more precise estimates of results and allow definitive determination of any economic benefit of multiple stage contacting. The importance of making such a determination can be appreciated by recognizing that carbon dosage requirements, predicted by the regression equations in Table 23, indicate that about 3 1/2 times more carbon is required for single-stage than for two-stage counter-current contacting. It is interesting to note that an example in Appendix A, for SCOD removal by physical adsorption, shows 3.3 times more carbon is required for single-stage than for two-stage counter-current contacting.

The cost of providing a second-stage carbon contactor would amount to about 0.39¢/1000 gal. This cost is for installed solids-contacting equipment designed at an overflow rate of 0.48 gpm/sq ft (see Table 17). The concentration of virgin powdered carbon, at 10¢/lb, which must be saved to justify a second-stage contactor, at a cost trade-off of 0.39¢/1000 gal., is 4.7 mg/ℓ. The concentration of regenerated carbon, estimated to cost 3¢/lb, which must be saved is 16 mg/ℓ.

Figure 50 shows predicted effluent SCOD versus carbon dosage for different feed SCOD concentrations as per the combined single-stage and two-stage counter-current regression equipment in Table 23. The insensitivity of effluent SCOD

FIGURE 50: CARBON TREATMENT: RELATIONSHIP BETWEEN FEED AND EFFLUENT SCOD AND CARBON DOSAGE



to carbon dosage is indicated by the slope of the lines which show that a 100 mg/ℓ incremental increase in carbon dosage results in removing only 1.8 mg/ℓ of additional SCOD. Comparison of carbon dosage required for different feed SCOD's at a given effluent SCOD value, indicate that an increase of about 24 mg/ℓ of carbon is required per mg/ℓ increase in feed SCOD.

As noted previously, the STOC removal data was not realistically described by an adsorption model. Combined single-stage and two-stage counter-current results did correlate fairly well when plotted as X/M vs C_o/M. A linear regression equation of $X/M = 0.045 + 0.78\ C_o/M$, with a correlation coefficient of 0.91 resulted. This observation lends credence to the applicability of this emperical relationship to the carbon contacting system studied.

Other Observations

As noted previously the variability of SCOD removal results precluded precise definition of the difference between single-stage and two-stage counter-current contacting effects. Variability during any given Run can be attributed to varying wastewater quality and plant operation. For example, carbon contactor feed and effluent SCOD concentrations varied from day to day as indicated by typical data in Figures 51 and 52. Some of the variation in effluent SCOD can be attributed to varying average daily carbon dosages. Comparison of carbon contactor effluent and feed SCOD concentrations at relatively constant carbon dosages indicated a general trend toward higher effluent concentrations at higher feed concentrations. However, no precise correlation was found.

Another possible source of variability might have been the presence of an unadsorbable fraction of SCOD. During Runs C-9 and C-11 an attempt was made to quantify any "unadsorbable" fraction by the following procedure. Aliquots of plant effluent composite samples were contacted with massive dosages (30-60 g/ℓ) of degassed virgin powdered carbon. Duplicate SCOD analysis were conducted on 0.45 micron membrane filtrate after one hour of contacting. Figure 53 shows results obtained during Run C-11. These results were not precisely modeled by either the Freundlich adsorption model or the

FIGURE 51: CARBON TREATMENT: TYPICAL EFFLUENT SCOD VARIATIONS

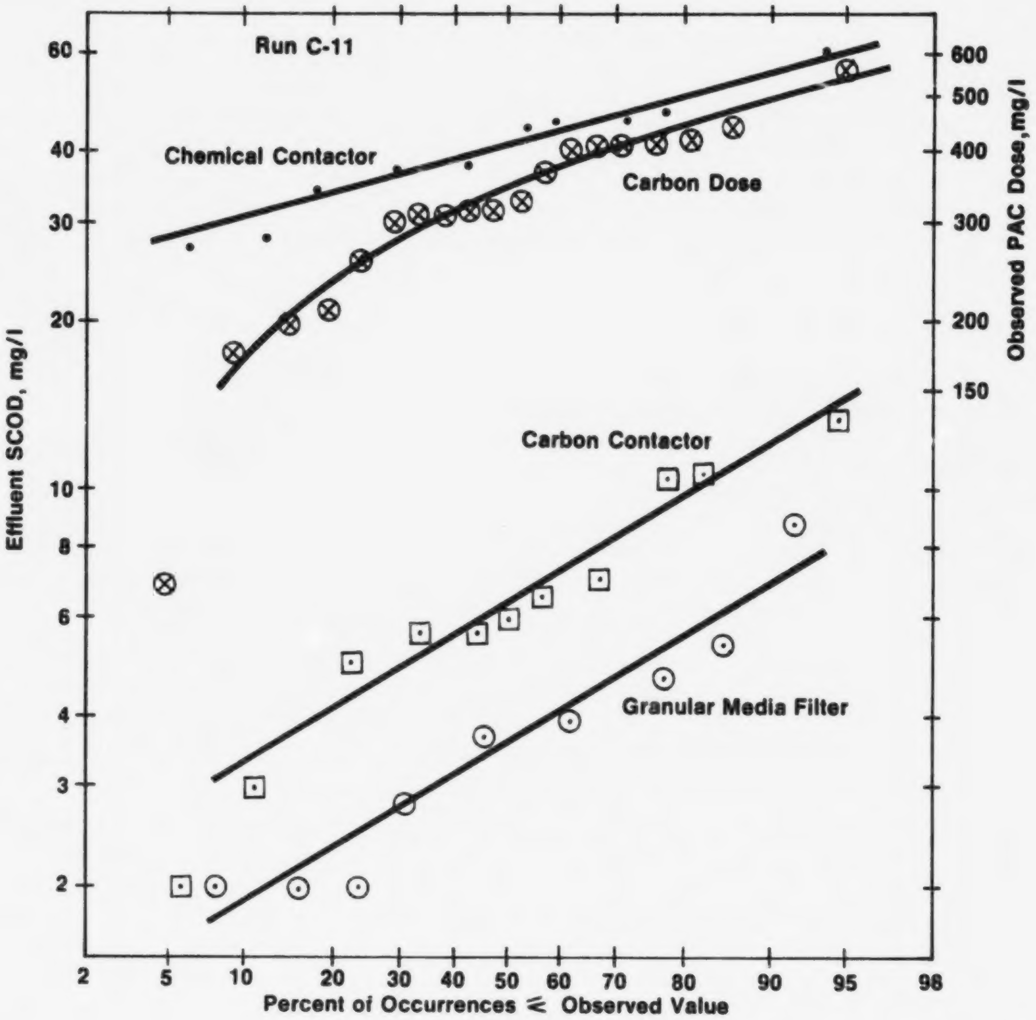


FIGURE 52: CARBON TREATMENT:
TYPICAL EFFLUENT SCOD VARIATIONS

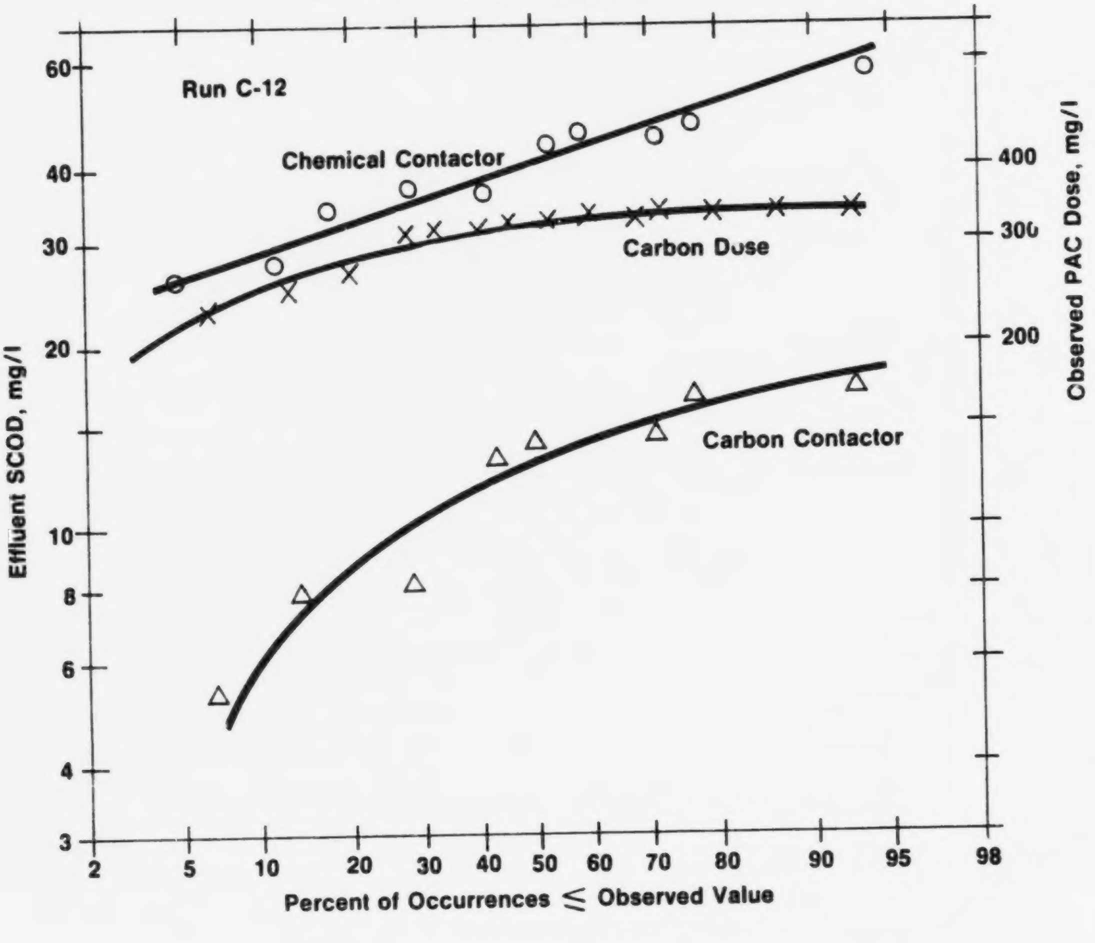
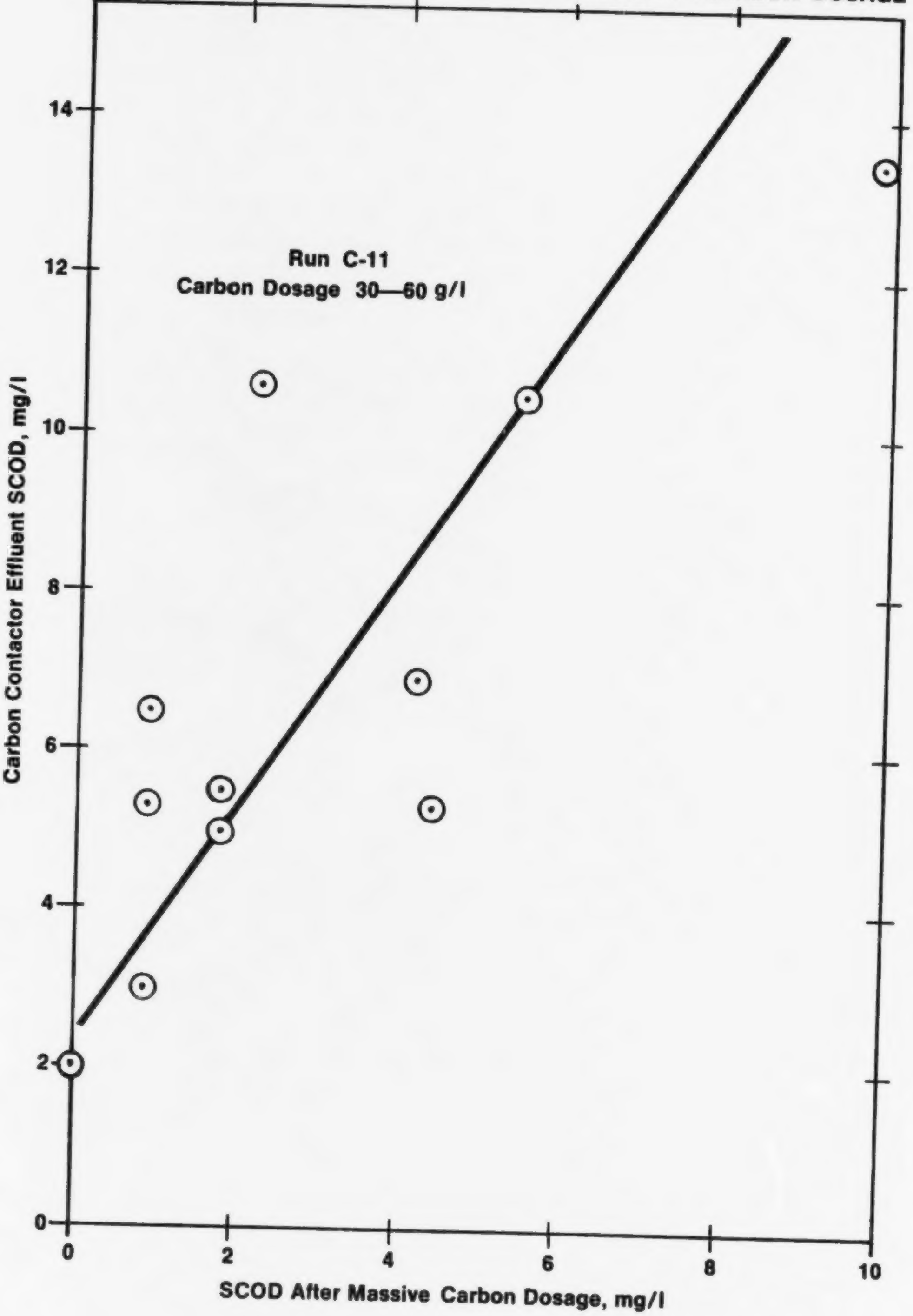


FIGURE 53: CARBON TREATMENT: EFFECT OF MASSIVE CARBON DOSAGE



X/M versus C_o/M model. The trend toward lower SCOD concentrations, after massive carbon treatment, at lower carbon contactor effluent SCOD would seem to indicate little if any SCOD was "unadsorbable".

Summary of Powdered Carbon Treatment Results

Solids-contact units were found to be very effective for use as powdered carbon contactor-clarifiers. The total solids recycle and variable area clarification zone features of the units used were considered to be key factors in providing a significant level of biological activity without odor problems and in accomplishing effective removal of carbon solids. Effective gravity clarification of carbon suspensions was achieved at overflow rates up to 0.8 gpm/sq ft without the use of flocculation aids.

Granular media filtration effectively removed carbon particles from carbon contactor effluent. However, the coal media size was too small to allow penetration and removal of suspended solids "in depth". Consequently, backwash recycle was excessive. Filtration cycles in excess of 24 hours should be avoided to minimize "mud ball" formation.

Soluble organics removal results were not precisely described by the Freundlich adsorption model. However, use of this model did indicate the presence of a soluble organic removal mechanism in addition to physical adsorption. Biological oxidation was the most probable mechanism. The more generally applicable emperical relationship, $X/M = K_1 + K_2\ (C_o/M)$, was found to precisely describe the pilot plant organic removal results (both SCOD and STOC). A statistical analysis of these results indicated that two-stage counter-current contacting was probably more efficient (i.e., required less carbon) than single-stage contacting. However, the data were too variable to precisely quantify the difference in carbon requirements. Additional pilot plant studies are strongly, recommended to more precisely define this difference. If differences of at least 20 to 30 mg/ℓ carbon dosages are found, then two-stage counter-current contacting would be an economic choice over single-stage contacting.

Both the Freundlich and the X/M versus C_o/M models indicate that the carbon dosage required to produce a constant system effluent SCOD, is linearly related to the carbon system feed SCOD. Therefore, if a reliable automatic feed SCOD analyzer (direct or indirect) were available, the carbon feed rate could presumably be automatically controlled to match the feed SCOD and thus, result in production of a uniform effluent SCOD. Such a device must measure soluble organics concentration very precisely since a small change in feed organics indicates a substantial change in carbon feed required to maintain a constant system effluent organics concentration. One possible approach entails use of ultra violet light absorption measurements to indirectly indicate soluble organic concentration.

CARBON REGENERATION SUBSYSTEM

Regeneration and reuse of spent carbon was to be evaluated during a six (6) month extension to the original 18 month research contract. Due to uncontrollable delays in funding and installation of carbon dewatering and regeneration equipment, only about two months of operating data was obtained. The carbon regeneration results should be considered very preliminary, but do provide a good basis for additional studies.

Due to the limited available time, the following approach was used to evaluate spent carbon regeneration and reuse. During carbon treatment Run C-11, spent carbon was allowed to build-up in the thickener, carbon contactor, inventory tank and holding tank. A total of about 4,500 pounds of virgin carbon was fed with some 1,200 pounds being advanced to the holding tank for vacuum filtration and thermal regeneration. After the first regeneration Run no virgin carbon was used, only regenerated carbon. The existing carbon inventory was regenerated and reused for the remainder of the study. This approach resulted in passing all carbon through the regeneration system at least once.

The following is a presentation and discussion of the carbon regeneration system consisting of graivty thickening, vacuum filtration, thermal regeneration and reuse.

Gravity Thickening

Normal operation of the thickener entailed intermittant feed and underflow withdrawal. Approximately 30-70 gallons of carbon contactor blowdown was fed to the thickener every 1/2 to 1 1/2 hours at a rate of 50 to 70 gpm. Thickener overflow was collected in a 55 gallon drum and pumped to the granular media filter backwash collection tank and subsequently recycled back to the carbon contactor. Thickener underflow was pumped to the spent carbon inventory tank, at about 5 gpm for 1/2 to 4 minutes durations each 10 minutes.

During operation of the regeneration furnace, thickener underflow had to be shut off because the single carbon inventory tank was used to inventory and sample regenerated carbon. Thus, just prior to a regeneration run, carbon concentration in the contactor and thickener was reduced by advancing carbon to the holding tank. During each regeneration Run, lasting about 20 hours, carbon was allowed to build up in the contactor and thickener. Thus, immediately after a regeneration Run, the flux of carbon (lb/day) through the thickener was considerably higher than normal.

The above operational procedures resulted in considerable variation of thickener solids loading and underflow concentrations. Table 24 shows the range of results. The average thickener loading for the 28 days of data presented in Table 24, was 7.8 lb/day-sq ft, and the weighted average of underflow solids 124 g/ℓ. Generally speaking, as thickener solids loading increases, the underflow solids concentration should decrease. Such a trend is not apparent from the results in Table 24. It is quite probable that the thickener was conservatively loaded.

Thickener feed and overflow solids were not routinely monitored. Therefore, the effect of thickener feed solids on thickener underflow solids concentration could not be determined. Carbon dosage and carbon contactor blowdown volume data in Tables 13 and 14 indicate that carbon blowdown solids were 30 to 50 g/ℓ.

TABLE 24: SUMMARY OF SPENT CARBON GRAVITY THICKENING RESULTS

Period	Spent Carbon Inventoried (lb)	Thickener Solids Loading[c] (lb/day-sq ft)	Thickener Solids Underflow Concentration (g/ℓ)
Prior to Run #1[a]	209	--	56
	599	7.6	151
	438	26.0	122
Average	1246	5.3	125[b]
Between Runs #1 and #2	381	--	121
	383	5.2	86
	249	9.8	52
Average	1013	7.1	91[b]
Between Runs #2 and #3	687	--	180
	615	18	134
Average	1302	14	158[b]
Between Runs #3 and #4	134	--	46
	469	8.1	121
	133	7.7	106
Average	736	9.8	105[b]

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[a] - Regeneration Runs

[b] - Weighted Average Based on Spent Carbon Advanced

[c] - Based on Underflow Solids

Summarizing the gravity thickening results, it appears that spent carbon solids of 30 to 50 g/ℓ can be thickened to 100 to 120 g/ℓ at a thickener solids loadings of 10 lb/day-sq ft or greater. It is recommended that future pilot plant tests be designed to evaluate the effect of loadings of at least 20 to 30 lb/day-sq ft and the effect of feed solids concentration on underflow concentration and to quantify the capture of carbon solids by the thickener.

Vacuum Filtration Dewatering

The vacuum filter was operated during each regeneration Run primarily to provide dewatered spent carbon for feeding to the fluidized bed furnace. No attempts were made to "optimize" filter operation.

Vacuum filter yields (capacity) were about three times greater than designed for. Thus, the filter was run for only one hour out of each 3 to 4 hours of furnace operating time. During each regeneration Run, the filter was operated for three to five one hour periods. Single grab samples were taken during each filter Run for determining filter feed solids concentration, filtrate solids concentration, filter cake solids concentration and filter cake thickness. The volume of carbon sludge dewatered and the polymer feed solution used during each Run was determined by inventory measurements (i.e., depth measurements in the carbon sludge holding tank and polymer feed tank). Because complete mixing conditions did not exist in the carbon holding tank, filter feed solids varied from Run to Run. Polymer dosages in the range of 0.1 to 0.7 lb polymer per 100 lb dry carbon solids were attempted by setting the polymer feed rate based on the known overall average holding tank solids concentration. The polymer used was Dow Chemical Company's C-31, a cationic polyelectrolyte with amine functional groups. Spent carbon sludge was stored in the holding tank for from 1 to 14 day periods prior to dewatering.

Table 25 presents grab sample results for each filter Run. Filter operating conditions were generally fixed by the type of machine used (i.e., a continuous belt vacuum filter). Actual drum submergence was less than the 33 percent design. The drum cycle times were

TABLE 25: SUMMARY OF SPENT CARBON VACUUM FILTRATION RESULTS

Regeneration Run #	1	1	1	1	1	2	2	2	2	2
Vacuum Filtration Run #	1	2	3	4	5	1	2	3	4	5
OPERATING CONDITIONS										
Cycle Time, min/rev	8.8	8.8	8.8	8.8	8.8	8.8	8.8	8.8	8.8	8.8
Submergence, %	--	--	22	27	30	--	--	28	28	28
Polymer Feed, % by wgt	0.3	0.4	0.4	0.2	0.2	--	0.8	0.8	2.5	0.2
Feed Solids, g /ℓ	132	115	111	105	105	148	97	71	26	262
Filtrate Solids, g /ℓ	--	8.4	6.3	19	18	7.9	6.1	6.8	15	19
Solids Capture, %	--	93	94	82	83	95	94	90	42	93
FILTER CAKE										
Thickness, inches	3/4	3/4	3/4	3/4	3/4	3/4	7/8	3/4	5/8	7/4
Moisture, % by wgt	62	79	77	78	75	80	78	58	77	78
Solids, lb /sq ft	1.8	1.0	1.1	0.9	1.1	1.1	1.1	1.8	0.6	2.2
Filter Yield, lb /hr-sq ft	12	6.8	7.5	6.1	7.5	7.5	7.5	12.2	4.1	15.0

TABLE 25 (cont.): SUMMARY OF SPENT CARBON VACUUM FILTRATION RESULTS

Regeneration Run #	3	3	3	3	4	4	4
Vacuum Filtration Run #	1	2	3	4	1	2	3
OPERATING CONDITIONS							
Cycle Time, min/rev	9.7	9.7	9.7	9.7	9.7	7.0	9.7
Submergence, %	28	29	28	28	27	24	27
Polymer Feed, % by wgt	0.5	0.4	0.9	2.4	0.1	<0.1	<0.1
Feed Solids, g /ℓ	162	109	91	29	90	70	106
Filtrate Solids, g /ℓ	7.5	21	36	--	45	17	--
Solids Capture, %	95	80	60	--	50	76	--
FILTER CAKE							
Thickness, inches	1	5/4	1	1	1	1/2	3/4
Moisture, % by wgt	78	78	79	79	79	78	78
Solids, lb /sq ft	1.5	1.7	1.4	1.1	1.4	0.63	1.0
Filter Yield, lb /hr-sq ft	9.2	10.5	8.9	6.8	8.9	6.1	9.8

arbitrarily chosen at about the maximum possible. Cake form times were about 6 and 12 percent of the cycle time. A dry time of about 44 percent of the cycle time was used.

Several preliminary, yet pertinent observations can be made concerning the data in Table 25 and other operational observations. First it was qualitatively established, very early in the test work, that polymer conditioning of carbon sludge was necessary to achieve a readily dischargable filter cake. A 3/4 to 1 inch thick filter cake could be formed and dried without polymer, but it did not discharge easily.

The level of polymer dosage appears to have effected carbon solids capture by the filter. At dosages of less than 0.2 lb C-31/100 lb dry carbon solids or at dosages greater than 0.9 lb C-31/100 lb carbon, solids capture was poor (generally less than 75 percent). Polymer dosages of 0.4 to 0.8 lb C-31/100 lb carbon results in from poor to fair captures of 80 to 95 percent.

The nature of the carbon solids results in filter cake shrinkage and formation of large "cracks" during cake drying. These cracks formed at about 1/4 to 1/3 of the dry cycle time and caused very low operating vacuums of less than about one and one-half inches of mercury during the dry cycle. Since the filter was equipped with a single vacuum pump and filtrate receiver, the form vacuum was also quite low, being in the range of one to five inches of mercury.

Because of varying operating parameters of cycle time, submergence and polymer dosage and varying filter cake thicknesses, the data in Table 25 does not show a precise correlation between filter yield and feed solids concentration. There is a general trend toward higher yields at higher feed solids concentration.

During all filter Runs, a total of 4300 lb of carbon sludge (dry weight basis) was fed to the vacuum filter during 16.2 hours of filter operating time. For the 28.3 sq ft filter, the overall average yield, assuming 90 percent capture, was 8.4 lb/hr-sq ft (e.g., 4300 x 0.90/16.2 x 28.3). The weighted average of filter feed solids was 12 percent solids, by weight.

Based on results of this study, a summary of suggested vacuum filter operation conditions and performance is presented in Table 26. An average filter cake density of 16 lb/cu ft was observed, excluding the obviously low moisture content samples of 62 and 58 percent in Table 25. Based on a 3/4 inch thick cake and a cycle time of 9 minutes, a filter yield of 6.7 lb/hr-sq ft was computed.

Future studies in at least three areas should result in demonstrating significantly improved filter performance. First, a tighter filter media should be evaluated for the purpose of improving solids capture. The media used was Polypropolene-907, which has an air flow rate classification of 300 CFM/sq ft, a 2/2 twill weave of 12 millimeter mono-filament and a thread count of 68 x 29. It is recommended that a similar type of media be used which has a lower air flow rate classification; for example, Polypropolene-873 which is rated at 30 CFM/sq ft. Use of the tighter media would probably result in reduced filter yields. Laboratory filter leaf test results indicated that solids captures in excess of 95 percent could be obtained with Polypropolene-873.

The second area of improving filter operation results involves more efficient use of the filter cycle time. From Table 26 it is seen that cake form time was only about 12 percent (1.1 min/9 min) of the cycle time. Had the full design submergence of 33 percent been used for cake form time, the cycle time could have been reduced to 3.3 minutes (9 x 12/33). The filter yield would have been expected to increase to over 18 lb/hr-sq ft (recall Equation (4) where yield $\propto$ 1/CT). Adequate dry time could be maintained at the reduced cycle time of 3.3 minutes.

The third area of improving vacuum filter performance would be the use of two vacuum receivers wherein form and dry vacuum could be at different levels. All other factors being equal an increase in form vacuum from 5 to 15 inches of Hg should result in 40 to 50 percent increase in yields.

Thermal Regeneration

The primary objectives of the fluidized bed carbon regeneration furnace tests were to determine the carbon losses across the system and demonstrate the operability of the system at pilot plant scale. The quality of regenerated carbon was evaluated by reuse in the carbon contactor treatment step.

TABLE 26:
SUMMARY OF SPENT CARBON DEWATERING DATA

Item	Value
Filter Cake Density	16 lb/cu ft
Filter Cake Thickness	3/4 inch
Dry Cake Weight (W)	1.0 lb/sq ft
Filter Drum Cycle Time (CT)	9 minutes
Filter Cake Form Time (FT)	1.1 minutes
Filter Cake Dry Time (D)	4.0 minutes
Filter Yield	6.7 lb/hr-sq ft
Filter Cake Moisture Content(MC)	78 percent
Drum Submergence	30 percent
Solids Capture	90 percent
Polymer Dosage (Dow C-31)	0.4-0.6 lb/100 lb carbon
Form and Dry Vacuum	5 inches of Hg

Because of limited time, evaluation of the effect of various furnace system operating variables was not attempted. The approach used was to operate the system as per the manufacturers and process consultants recommendations. Valuable input on furnace operation was received from Mr. Ed Berg of the Environmental Protection Agency and Mr. Bob Thompson of Stamford, Connecticut (a private consultant).

Four batch regeneration Runs were completed. A fifth was started but equipment failures caused termination of that Run. The following is a discussion of each separate Run in the order that they were conducted.

Regeneration Run #1

The purpose of the first Run was to check out operating and testing techniques and procedures, and obtain preliminary carbon loss data. Furnace system operating conditions are presented in Table 27, which summarizes operating data for all four regeneration Runs.

During Run #1, the oxygen analyzer was not functioning properly and control of furnace exit oxygen content was impossible. Normal operation would have entailed adjusting bed injection gas flow rate to provide about 1 1/2 percent excess oxygen prior to carbon feed and about 1/2 percent after carbon feed. Because of the inoperative oxygen meter, the air to fuel volume ratio was maintained well below the theoretical stoichiometric ratio of 10:1. Since the actual ratio was 6.7:1 (air:gas), it was presumed that no oxygen existed in the furnace off-gases.

From Table 27 it is seen that the temperature at the top of the furnace freeboard was 70°F higher than the sand bed temperature. This observation indicates that heat was being generated between the top of the sand bed and the freeboard thermocouple. Burning of either gas, carbon or both in this region obviously occurred. Carbon loss results indicate that considerable carbon was burned. During Run #1, 1250 pounds of total solids was advanced from the spent carbon holding tank to the vacuum filter station. An estimated 950 pounds was fed to the furnace. The estimated difference, or 300 pounds of solids, left the vacuum filter station via filtrate (120 lb), belt washing and spillage (80 lb) and final

TABLE 27: CARBON REGENERATION: FURNACE OPERATING CONDITIONS

OPERATING CONDITIONS:	1	2a	2b	3	4
Burner Air Flow, SCFM	126	65	125	126	119
Burner Gas Flow, SCFM	6.7	2.8	5.1	6.2	5.1
Bed Injection Gas Flow, SCFM	12.0	8.3	11.4	9.8	9.2
Off-Gas Oxygen Concentration, % by Volume	--	<2.0	<0.8	0.0	0.0
Temperature Profile (°F)					
Freeboard	1600	1740	1650	1650-1750	1510
Sand Bed	1530	1740	1650	1650-1750	1540
Fire Box	1920	1920	1890	1900	1940
Pressure Profile (inches of H_2O)					
Freeboard	-4	-6	-4	-4	-4
Bottom of Sand Bed	28	22	23	49	46
Across 24 inches of Sand Bed	28	25	26	40	33
Fire Box	42	26	40	67	63
Scrubber and Quench Water:					
Flow, gpm	62	60	60	66	57
Temperature, °F	61	51	61	62	56
Furnace Operating Time, Hours	18	4	21	22	12

cleaning of piping, flocculation tank, vacuum filter vat and furnace feed hopper (100 lb). Of the estimated 950 pounds of solids fed to the furnace only 350 pounds were recovered in the scrubber system decant tank. Thus, 63 percent of the total solids fed were lost. The recovered carbon was inventoried and advanced to the carbon contactor feed system.

The furnace temperature and pressure profiles were very stable during the course of the Run. No evidence of carbon was observed in the furnace stack gas. The automated carbon feed system worked well, with an average of about 53 pounds of dry solids being fed per hour to the furnace (950 lb /18 hr). No obvious explanation was apparent for the very high carbon losses.

Regeneration Run #2

Run #2 was started because of a need for regenerated carbon in the pilot plant. As the furnace was brought up to operating temperature, considerable difficulty was experienced in trimming the stack-gas oxygen level to less than 2 percent by volume. As indicated in Table 27, the air flow rate was considerably reduced during Run #2a. During furnace start-up it was observed that as bed injection gas flow was reduced, the freeboard temperature was also reduced, compared to the sand bed temperature. Unfortunately as bed injection gas flow was reduced, the stack-gas oxygen concentration increased above 2 percent by volume.

Run #2a was commenced even though the air and gas flow rate was slightly less than the design fluidization rates. Within a four hour period, bed fluidization was lost twice. Pressure readings became very erratic and the upper and lower sand bed temperature differed significantly. It was decided to terminate the Run, leaving the regenerated carbon and spent carbon inventories in place. A representative of the furnace manufacturer and the process consultant (Mr. B. Thompson) were summoned to assist in operation and evaluation of the furnace system. Under their supervision, Run 2b (continuation of 2a) was completed at the operating conditions indicated in Table 27.

Of 1010 pounds of carbon solids fed to the vacuum filter station, some 760 pounds were estimated to have reached the furnace. Only 400 pounds of solids were recovered, as determined from inventory measurements. Thus, carbon losses were at least 47 percent. The furnace feed solids were 4.7 percent ash and the furnace product 23 percent ash. Visual inspection of the regenerated carbon indicated the presence of fine sand particles. As with Run #1, no obvious explanation for gross carbon losses was apparent for Runs 2a and 2b. The process and equipment consultants did offer several suggestions for improving equipment performance. For example, the furnace shell was pressure tested and several major air leaks found. It was conjectuered that leaks around the top of the furnace (at a flange) might have allowed air to be drawn into the unit due to the slight vacuum conditions at this location. Thus, all leaks were sealed by welding prior to Run #3.

It was also recommended that an additional two feet of sand be placed in the furnace to provide a greater detention time of carbon within the bed. This was also done prior to regeneration Run #3.

Regeneration Run #3

At this point in the test work most minor equipment and operating technique problems had been repaired or satisfactorily worked out. Some new plumbing and procedures were also implemented to more precisely monitor the amount of carbon reaching the furnace. In essence, all carbon bearing flows from the vacuum filter station, except the cake discharge, were collected and inventoried for volume and concentration of total suspended solids. This included any spillage and hosing down of the filter station after use.

During Run #3 the furnace operating conditions were similar to previous Runs as indicated in Table 27. The increased sand bed depth did result in higher pressure readings. The freeboard temperature was, once again, equal to the sand bed temperature. During the Run, the sand bed temperature set point (automatic carbon feed controller) was increased by 25°F increments at about one-hour intervals. The air and gas flow rates were not changed. To achieve the higher bed temperature the carbon feed rate was automatically reduced. Immediately after each adjustment of bed temperature set point, the freeboard temperature dropped about 30 to 50°F. However, within one-half hour, the freeboard and sand bed temperatures would merge to the same values.

Carbon losses were excessive once again. Only 520 pounds of total solids were recovered from 1100 pounds fed to the furnace, for a loss of 53 percent solids. The furnace feed (spent carbon) ash content was 4.7 percent by weight and furnace product ash content 30 percent by weight. As in Run #2 considerable sand was present in the furnace product. About 65 percent of the non-ash material was lost across the furnace. The average furnace feed rate during Run #3 was 50 pounds of total dry solids per hour (1100 lb/22 hr).

Regeneration Run #4

Run #4 was started without a clear cut direction in mind to reduce carbon losses. The only change in furnace operating conditions was to set the automatic carbon feed controller to maintain a sand bed temperature of 1540°F which was 100 to 200°F less than for the two previous Runs. The stack off-gas oxygen level was again mantained at zero. During Run #4, the furnace system operated beautifully. No adjustments of any operating variables were made during the Run. The furnace temperature profile remained absolutely constant during the 12 hour Run. For the first time, the freeboard temperature remained slightly below (30°F) the sand bed temperature.

Some 740 pounds of total solids were fed to the vacuum filter, of which 360 pounds did not reach the furnace. This high percentage loss of carbon solids across the vacuum filter station was presumably partly due to the very poor captures experienced (see Table 24). In addition, the 750 pounds started with was considerably less than the 1010 to 1300 pounds in previous Runs. Consequently, spillage and clean up losses were proportionally higher.

Of the 380 pounds of total solids (740 minus 360) fed to the furnace 320 pounds were recovered. This resulted in total solids losses of 16 percent. The ash content of the furnace feed was 11 percent and of the furnace product 28 percent. Apparently some previously regenerated product had been cycled through the carbon contacting step and advanced to the spent carbon holding tank. The increase in spent carbon ash content from 4.7 percent in Runs #2 and #3 (basically once used carbon) to 11 percent in Run #4 indicates recycling of regenerated carbon. Small gold-colored flecks routinely seen in the furnace product were visually observed in small quantities in Run #4 furnace feed material.

Regeneration Run #5

With only enough time left for one additional furnace Run, a special test was designed to demonstrate that low carbon losses could be obtained with the fluidized bed furnace system. It was decided to operate the furnace without bed injection gas. The oxygen level was controlled by adjustment of burner air and gas flows. Thus, no excess oxygen would enter the sand bed region. It was recognized that this mode of operation would result in excessive temperature (about 3000°F) in the fire chamber of the furnace. The furnace manufacturer was consulted about probable damage to the furnace. It was indicated that the stainless steel tyeres would probably be deformed at 3000°F but that they could be replaced if damaged.

Run #5 was started with a burner air flow of 118 SCFM and a burner gas flow of 4 SCFM. The stack off-gas oxygen content was under 1/2 percent by volumes. The carbon feed temperature set point was at 1400°F. After two hours of furnace operating time, the fire box refactory failed (melted), and Run #5 had to be aborted. No data could be collected to determine carbon losses. A re-evaluation of furnace refactory material characteristics indicated that the fire box refactory was not designed for temperature in excess of about 2250°F. Three of the seven tyeres sustained severe damage also, with one being completely melted.

Discussion:

The preliminary carbon regeneration test results indicated gross burning of powdered carbon. The 83 percent recovery obtained in Run #4 was encouraging but still not as high as the 90 percent plus anticipated by the developers of the process. Obviously, there was a major shortcoming in the furnace design, or possibly, operation. Review of the furnace design revealed a weakness. The carbon feed point was located about 6 inches above and

16 degrees horizontally from one of the sand bed gas injection nozzles. Thus, spent carbon entered the fluidized bed at a point only 8 inches from a gas injection nozzle. Apparently, excess oxygen was not being consumed by preferential burning of bed injection gas. It superficially seemed desirable to feed carbon at a higher elevation in the sand bed, further removed from the gas injection points. Actually, two new carbon feed points were installed 18 and 36 inches above one of the six (6) gas injection nozzles. Lack of available time precluded determing the effect of these new carbon feed points.

A similar study of regeneration of powdered carbon by a fluidized bed furnace was conducted by others shortly after the end of this study. In this other study, a similar sized furnace was used without bed injection gas. Cooled off-gas (void of oxygen) was recycled to control furnace operating temperature. Preliminary results from operation of this furnace indicated carbon recoveries well in excess of 90 percent of the carbon[33]. This fact coupled with acceptable results from a study using a 4 lb/hr pilot plant[1] indicate that the fluidized bed furnace approach to powdered carbon regeneration is feasible. It is, however, quite obvious that additional development studies of the furnace used in the present study are required before it can be considered a potential alternate to other powdered carbon regeneration systems.

Reuse of Regenerated Carbon

Over a two week period about 1600 pounds of once regenerated carbon was used in the pilot plant. Chemical pretreatment was with lime at a treatment pH of 10.9. Approximately 0.77 MG of chemically treated wastewater was processed by single-stage carbon contacting. Table 28 presents the pertinent results for this carbon treatment Run (C-15) and, for comparison, the data from the previous carbon Run (C-11) using virgin carbon. The overall average carbon dosage shown is about 250 mg/ℓ, (i.e., 1600 lb/0.77 MG equals 250 mg/ℓ). However. the actual average daily carbon feed rate varied over a range of 150 to 300 mg/ℓ.

The organic loadings (C_o/M) for Runs C-11 and C-15 were 0.12 and 0.13 mg/ℓ feed SCOD per mg/ℓ carbon. The organic removals (X/M) were 0.10 and 0.098 mg/ℓ SCOD removed per mg/ℓ carbon, respectively. Considering the similar effluent

TABLE 28 : COMPARISON OF REGENERATED AND VIRGIN CARBON TREATMENT EFFECTIVENESS

	Virgin Carbon		Regenerated Carbon	
Run Number	C-11		C-15	
Length of Run, Days	20		14	
OPERATING CONDITIONS:				
Hydraulic Loading, gpm/sq ft (Carbon Contactor/GMF)	0.75/2-5		0.76/1.3	
Carbon Dosage, mg/ℓ	345		~250	
PROCESS TREATMENT RESULTS:				
Effluent From	Chem. Stage	Carbon Cont.	Chem. Stage	Carbon Cont.
Suspended Solids, mg/ℓ	7.7	53	14	35
SCOD, mg/ℓ	41	6.4	33	8.4
Temperature, °C	22	--	20	--

SCOD's of 6.4 and 8.4 mg/ℓ, it is apparent that no gross difference existed between virgin and once regenerated carbon, based on treatment effectiveness. These results are obviously of limited scope and future studies are required to more precisely define the properties and treatment effectiveness of regenerated carbon.

TYPICAL PILOT PLANT EFFLUENT QUALITY

Figures 54 through 58 present average pilot plant effluent quality for five different combinations of chemical and carbon treatment conditions. Average raw wastewater quality values are also noted. It should be noted that average results are presented and not median values shown in previous tabulations. Removal of COD ranged from 87 to 97 percent. Suspended solids and total phosphorus removals range from 94 to 98 percent and 91 to 97 percent respectively. Overall performance of the PAC-PCT system was considered quite good and variations in effluent quality during any given Run not unreasonable. Recall that the pilot plant was operated under relatively constant chemical and carbon dosage conditions. Varying raw wastewater quality (especially SCOD and phosphorus) account for some of the variations in plant effluent quality. If continuous on-line evaluation of SCOD and phosphorus were accomplished, then chemical and carbon feed rates could probable be adjusted to maintain a more uniform plant effluent quality. Future studies should be designed to evaluate such an operating approach.

The PAC-PCT process studied did not incorporate a nitrogen removal scheme. Since the entire plant flow was void of oxygen, little or no nitrite or nitrate nitrogen should have been present. Any nitrogen material would have been present principally as ammonia or organic nitrogen. After carbon treatment and removal of suspended solids the major form of nitrogen was probably ammonia. Table 29 shows several single composite ammonia nitrogen profiles through the pilot plant. The data indicates that a slight increase in ammonia nitrogen occurred through the pilot plant, presumably due to the breakdown of organic nitrogen.

FIGURE 54: PILOT PLANT EFFLUENT QUALITY FOR RUNS F-3 AND C-14

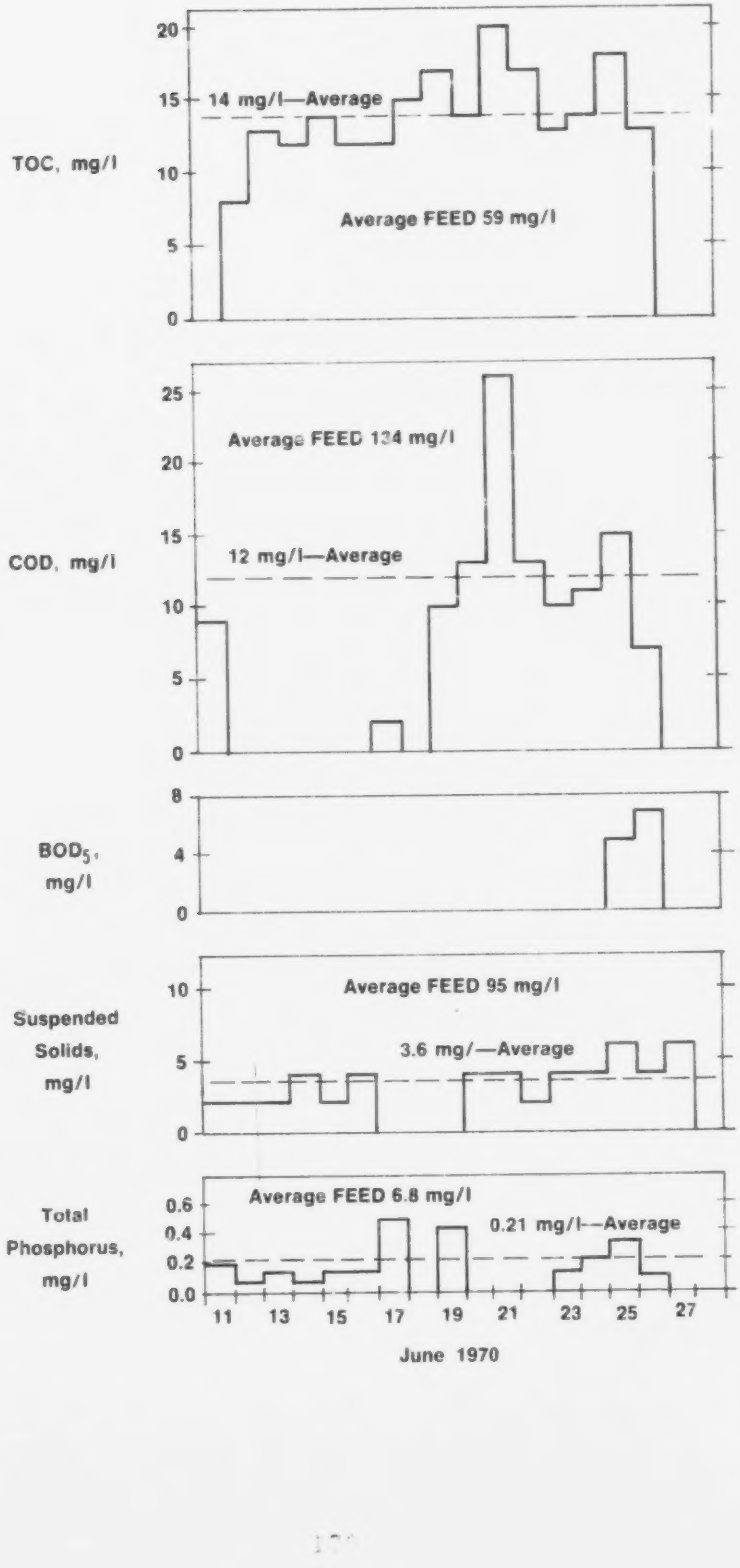

FIGURE 55: GRANULAR MEDIA FILTRATION:
FOR RUNS A-2 AND C-7

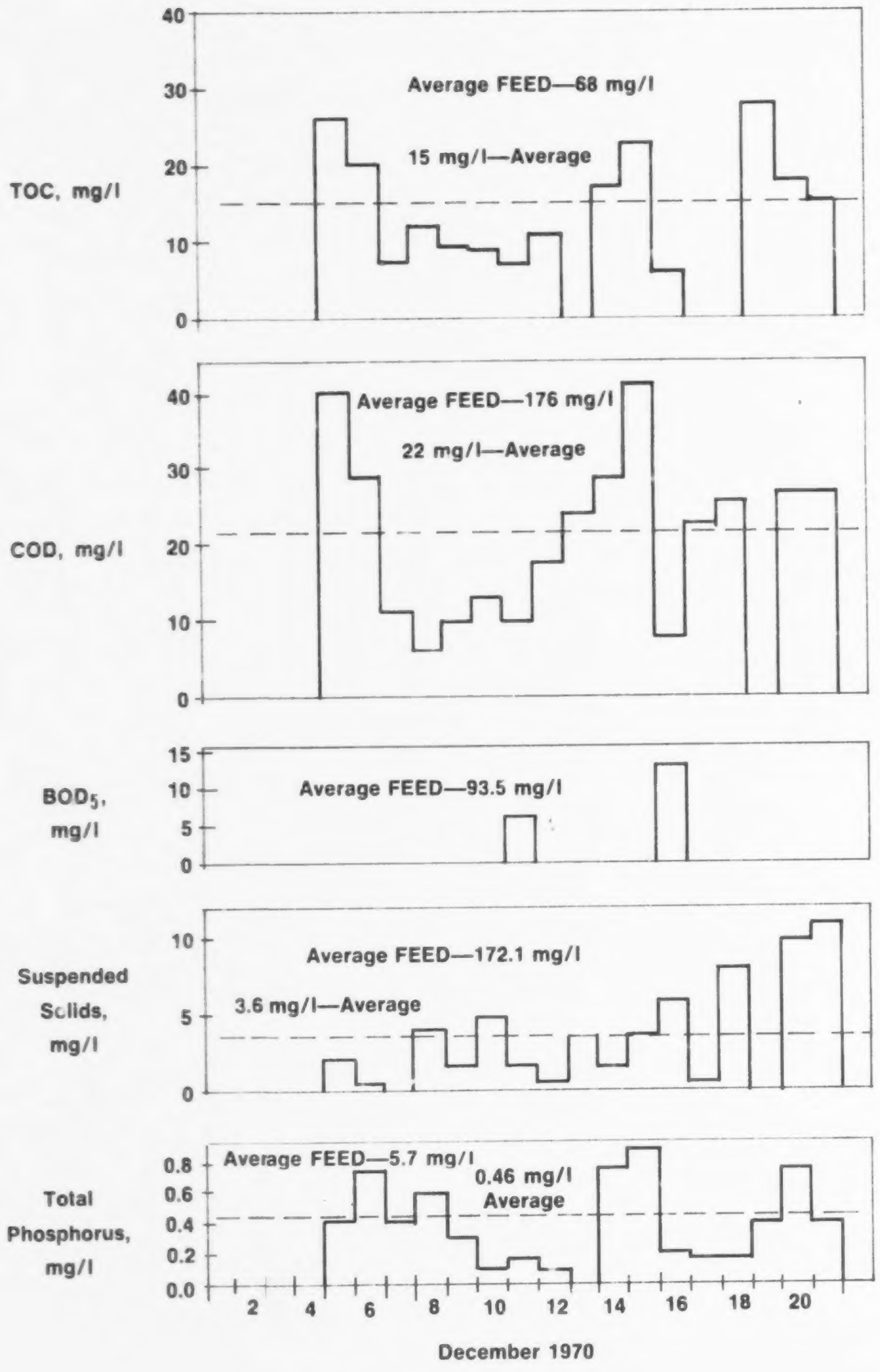

FIGURE 56: PILOT PLANT EFFLUENT QUALITY
FOR RUNS L-9 AND C-8

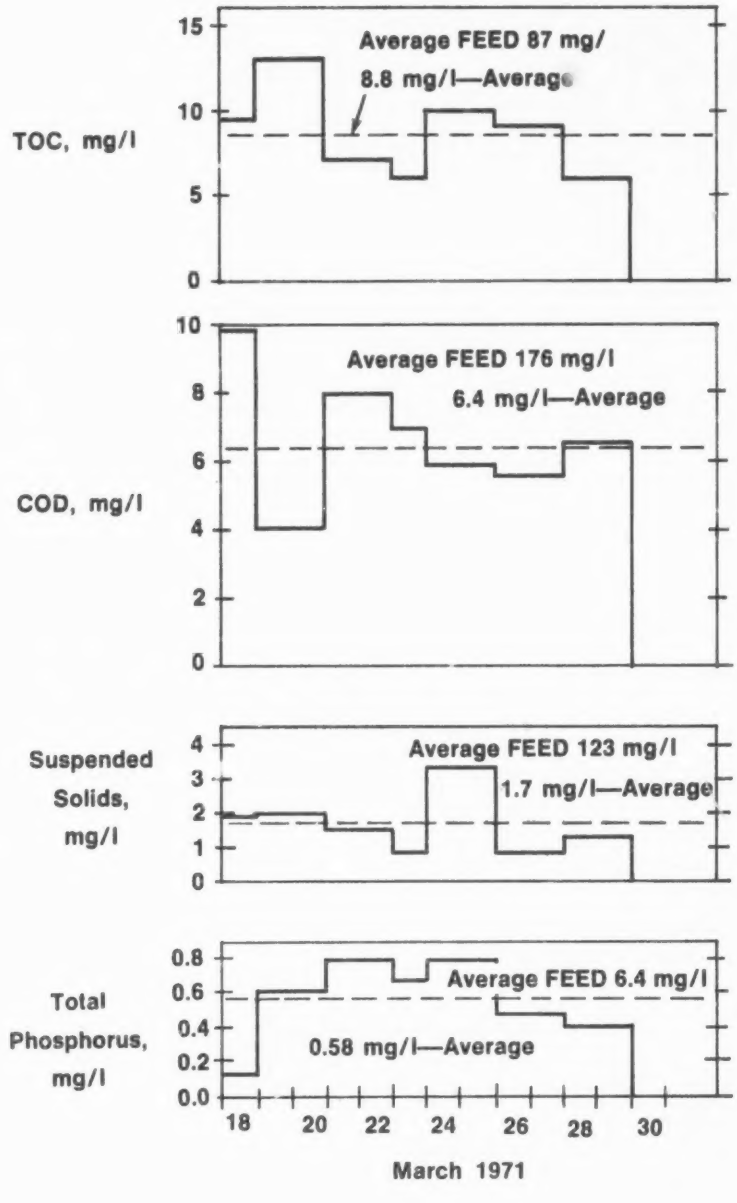

FIGURE 57: PILOT PLANT EFFLUENT QUALITY FOR RUNS L-11 AND C-9

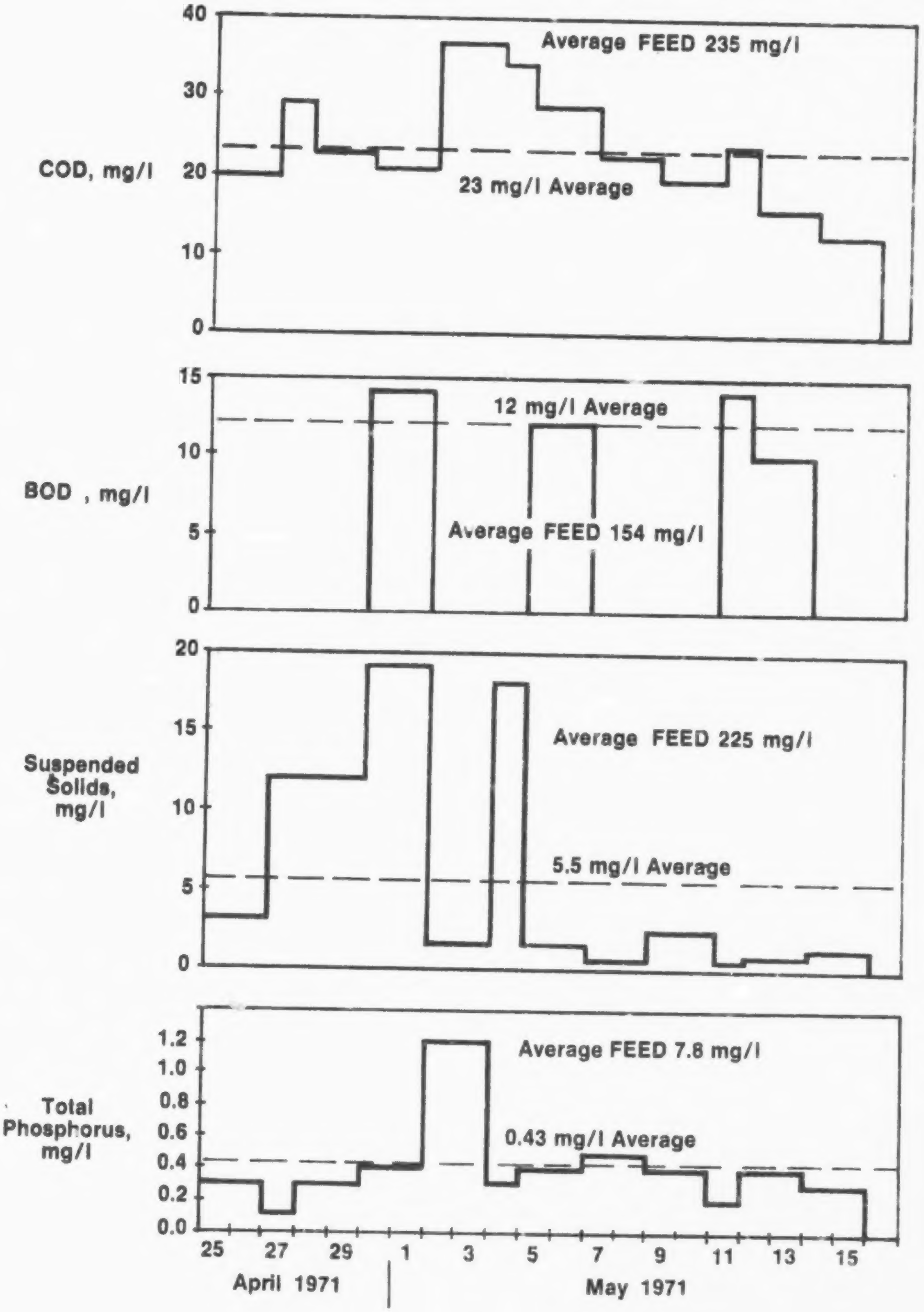

FIGURE 58: PILOT PLANT EFFLUENT QUALITY
FOR RUNS L-4 AND C-11

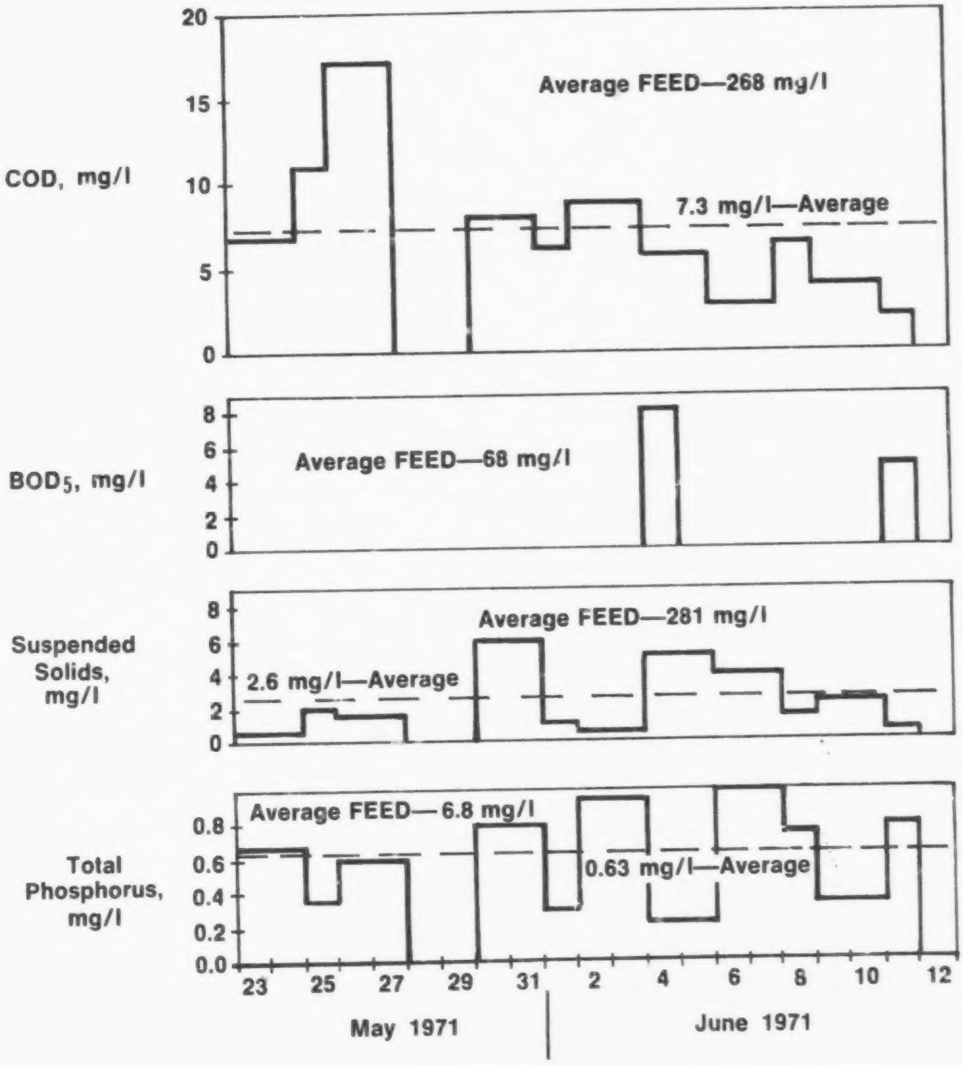

TABLE 29: AMMONIA NITROGEN PROFILE, mg/ℓ AS N

Run #	Raw Sewage	Chemical Effluent	1st PAC Stage Effluent	2nd PAC Stage Effluent	Plant (Filter) Effluent
C-1	8.8	7.4	8.6	9.0	9.4
C-2	7.2	10.1	9.9	10.1	10.7
C-3	10.6	11.0	11.3	12.0	11.1
C-5	8.5	9.0	9.6	--	10.2
C-7	9.5	10.9	--	10.9	11.0
AVERAGE	8.9	9.7	--	--	10.5

COMPARISON OF LABORATORY AND PILOT PLANT RESULTS

A direct comparison of laboratory and pilot plant treatment results is impossible since the laboratory study was conducted prior to pilot plant operation. However, certain pilot plant results were qualitatively predicted by laboratory tests.

The removal of SCOD by alum treatment was predicted by jar tests. The insolubilization of phosphorus by alum, ferric chloride and lime treatment was also fairly well predicted. The relative effect of lime dosage, or more specifically treatment pH, on hardness distribution and sludge production was also indicated by jar tests.

Chemical-sewage floc settling rates were not precisely predicted by jar tests. For example, jar test data in Figure 9 show that, without polyelectrolyte, lime-sewage floc settling rate of 1.4 inches per minute, which is equivalent to 0.8 gpm/sq ft was observed. Pilot plant results in Table 9 show that a well clarified effluent was produced at a hydraulic loading of 1.3 gpm/sq ft, for similar treatment pH's. The relative effect of polyelectrolyte on chemical-sewage floc settling rates was predicted. That is, the use of polyelectrolyte did result in providing effective clarification at increased hydraulic loadings.

Effective clarification of carbon slurry at a carbon contactor hydraulic loading of 0.8 gpm/sq ft was not predicted by laboratory jar test results. However, the laboratory tests did indicate a "self flocculating" property of concentrated carbon slurry. Being able to maintain a concentrated carbon slurry in the carbon contactors could explain the effective clarification obtained.

Laboratory equilibrium adsorption isotherm tests were not indicative of the organic removals obtained by the pilot plant. The obvious shortcoming of these laboratory tests was the inability to account for organic removal by any mechanism other than physical adsorption. The usefulness of the laboratory equilibrium isotherm test for predicting organic removals is seriously questioned.

In summary, certain laboratory tests did provide a "first approximation" of pilot plant treatment results. It is quite safe to note, however, that reliable design and efficiency data can only be obtained by pilot or full scale plant operation over an extended period of time.

OPERATIONAL AND EQUIPMENT PROBLEMS

Chemical Treatment System: Difficulty was experienced with the dry lime [$Ca(OH)_2$] feeder system during humid and rainy periods. The lime adsorbed moisture, balled up and bridged in the feeder. Though several treatment upsets were experienced, due to reductions of treatment pH, major problems were averted by operator attention. Basically, the operators would have to remove or crush balled up lime which blocked the two lime feeder discharge ports. These lime feedings problems could have been averted had lime been stored in a dry place.

Difficulty was experienced in maintaining a uniform effluent pH with the automated acid neutralization system due to the following reasons:

1. Insensitive acid feed pumping equipment.
2. Inadequate acid mixing and pH sensing, resulting in erratic signals to the automated acid control system.
3. Frequent interruption of flow due to timer controlled chemical sludge blowdown.

Two features of the existing acid neutralization system were considered poor design: a) feeding acid in the effluent launder, and b) off-on acid feed pump operation. It is recommended that acid be fed to a tank with 1 to 2 minute detention time which is completely mixed. In addition, off-on acid feed pump should be eliminated and an automatically controlled variable feed pump used. On two occasions, the acid feed line (1/2 inch diameter black iron pipe) was completely plugged with a fairly hard chemical deposit. It was cleared by rodding and flushing.

The solids-contact chemical treatment unit performed very well considering that it was basically designed as a potable or industrial water treatment unit.

Twice during this study, after about three months of operation with lime treatment, the sludge circulation pumping turbine clogged with a mixture of rags and $CaCO_3$ deposits. The deposits were easily removed by hand scraping.

This problem, a loss of mixing turbine pumping capacity, should be overcome by redesign of the turbine blades to prevent rag build-up.

No significant $CaCO_3$ deposits, other than those noted above, were observed in the chemical treatment unit during more than six months of lime treatment operation. Less than about 3/16 inch deposition of $CaCO_3$ was observed on the interior walls of the reaction zone and no significant deposition was ever found on the walls of the clarification zone.

After ten months of operation, the rubber lining in an eccentric sludge blowdown valve wore out and the valve internals had to be replaced. This failure was presumably due to the presence of grit in the raw wastewater which caused erosion of the rubber lining.

PAC Treatment System: No problems were encountered with the solids-contact units. Powdered carbon dusting problems originally encountered during carbon make-up were solved with the assistance of a carbon manufacturer. Covered make-up and feed tanks, a vacuum collection and a water scrubbing recirculation system, and careful handling and mixing (wetting) minimized carbon dusting.

Several corrosion and/or erosion problems were encountered in the concentrated PAC slurry handling system. Cast iron and steel components of pumps and pipe fittings corroded badly. In one instance, a stainless steel high speed (1750 rpm) mixing impeller shaft corroded at the water line in the virgin PAC make-up tank resulting in a shear failure at that point. A diaphragm pump with stainless steel ball checks was found suitable for pumping virgin PAC. For transferring solids-contact unit or thickener underflow, peristaltic pumps with rubber tubing were found adequate.

Granular Media Filtration System: Several structural and hydraulic problems were experienced with the 3.5 ft diameter plexiglass filter and appurtenances. A significant process problem was associated with obtaining effective backwashing of the filter bed. Some of the causes were:

(1) unsatisfactory backwash water distribution, (2) insufficient bed expansion during water backwash (less than 20 percent expansion at about 20 gpm/sq ft) and (3) "mud ball" formation during long filter Runs (50 hours). The first problem was effectively reduced by installation of a stainless steel underdrain septum over the entire filter bed area. The second problem was eliminated by installation of a larger backwash pump and larger piping. This allowed attainment of 25 percent filter bed expansion at a 26 gpm/sq ft backwash rate. The third problem was eliminated by backwashing the filter at least once each day.

Process Stream Sampling: Considerable difficulty was experienced throughout this study in obtaining a composite sample of raw wastewater. Daily cleaning and frequent operator surveillance of the sampler shown in Figure 24 was necessary in order to collect a representative 24-hour composite sample. A check of the sampler performance was made by comparing the suspended solids content in an automatically collected composite sample and a 24-hour composite of hourly grab samples taken from the raw wastewater sump. Less than 5 percent difference was found.

The possibility was considered that carbon present in the carbon contactor effluent 24-hour composite samples adsorbed additional SCOD due to the extended contact period. This possibility was evaluated by collection of two identical 24-hour composites of hourly grab samples. One grab sample was immediately filtered through a 0.45 micron membrane whereas the other was stored as collected. Soluble COD analysis of the two composite samples indicated that the presence of 30 mg/ℓ of spent carbon caused no discernable reduction in SCOD.

Instrumentation: Problems were encountered with nearly all plant instruments. Most notable were the electronic devices (recorders and controllers) which exhibited considerable corrosion of electrical contacts. The relatively humid plant atmosphere, the presence of ferric chloride fumes and lime dust were the obvious cause of corrosion. Obviously, delicate mechanical and electrical control devices and instruments should not be subjected to such an atmosphere for extended periods of time.

Considerable difficulty was experienced in maintaining the laboratory TOC analyzer in operable condition. Nearly 20 percent of the original instrument cost was spent in the first year for instrument calibration and repairs.

Carbon Regeneration System: The only significant problem encountered with the vacuum filtration operation was the variable feed solids concentration. The cause was the lack of well mixed sludge in the spent carbon holding tank. This problem of nonuniform solids concentration made it impossible to achieve the desired polymer conditioning chemical dosage. Provisions for better mixing of vacuum filter feed sludge would eliminate this problem.

Numerous problems were experienced with the fluidized bed furnace during start-up and during the first two regeneration Runs. The problem of gas leaks in the steel furnace shell has already been alluded to. The control valves on the two gas and single air flow lines were very insensitive and made precise adjustment of flow difficult. An exposed hot (1500°F) off-gas duct created a fire hazard and had to be insulated.

On several occasions, the dewatered carbon feed line plugged at the furnace inlet, presumably due to drying of carbon sludge at this point. Had the carbon feed inlet been water-jacketed, it is probable that this clogging problem would have been minimized. There was no isolation valve on the carbon feed inlet pipe. Thus, when plugging occurred, the furnace air and gas flow had to be reduced to less than that required for bed fluidization. Otherwise, hot sand would fall out of the carbon feed line when unplugged. A suitable isolation valve and quick-couple connection was provided by the furnace manufacturer upon request.

In general, the furnace and off-gas scrubbing system performed with a minimum of operational problems.

SECTION VIII

ECONOMIC ANALYSIS

The following economic analysis of the PAC-PCT process evaluated during this study is presented for the purpose of showing the relative economic impact of different types of chemical treatments and powdered carbon dosages. The economic analysis is presented in three parts: chemical treatment costs, carbon treatment costs and total treatment costs. The analysis is based on the treatment of Salt Lake City raw municipal wastewater. The average design flow used for analysis was 10 mgd and peak design flow was 15 mgd. All unit operations employed are presented in Figure 59, which shows the overall PAC-PCT system. Assumed unit costs are presented in Table 30. Major equipment sizes are presented in Table 31.

CHEMICAL TREATMENT COSTS

Estimated chemical treatment costs are shown in Table 32. The process design parameters used to size equipment and determine chemical costs are those presented in Table 12 (page 103). Predicted effluent quality for all three chemical treatments is 5-10 JTU turbidity, 10-25 mg/ℓ SS, 0.5 to 1.0 mg/ℓ total phosphorus and 55-60 mg/ℓ COD.

Preliminary treatment includes bar screening and comminution. Incineration operation and maintenance (O&M) costs include operators, utilities (power, water and fuel), maintenance and haulage of ash to land fill disposal. The power and maintenance and the supervision and labor costs shown are for one-half of the estimated total PAC-PCT treatment plant costs, excluding incineration O&M costs. The contingency costs (50 percent of total capital costs) include engineering, administration and legal costs in addition to interconnecting piping, valves, instruments, buildings, etc.

The absolute costs shown in Table 32 are considered reasonable and the relative costs for different chemical treatments valid. It is seen that alum treatment is the least expensive chemical treatment approach. Capital costs for the three types of chemical treatments are approximately the same. The major cost differences are operational costs for chemical and, in the case of lime treatment, incineration operation and maintenance (O&M) costs. The incineration O&M cost difference is due to

FIGURE 59:

POWDERED ACTIVATED CARBON PCT TREATMENT SYSTEM

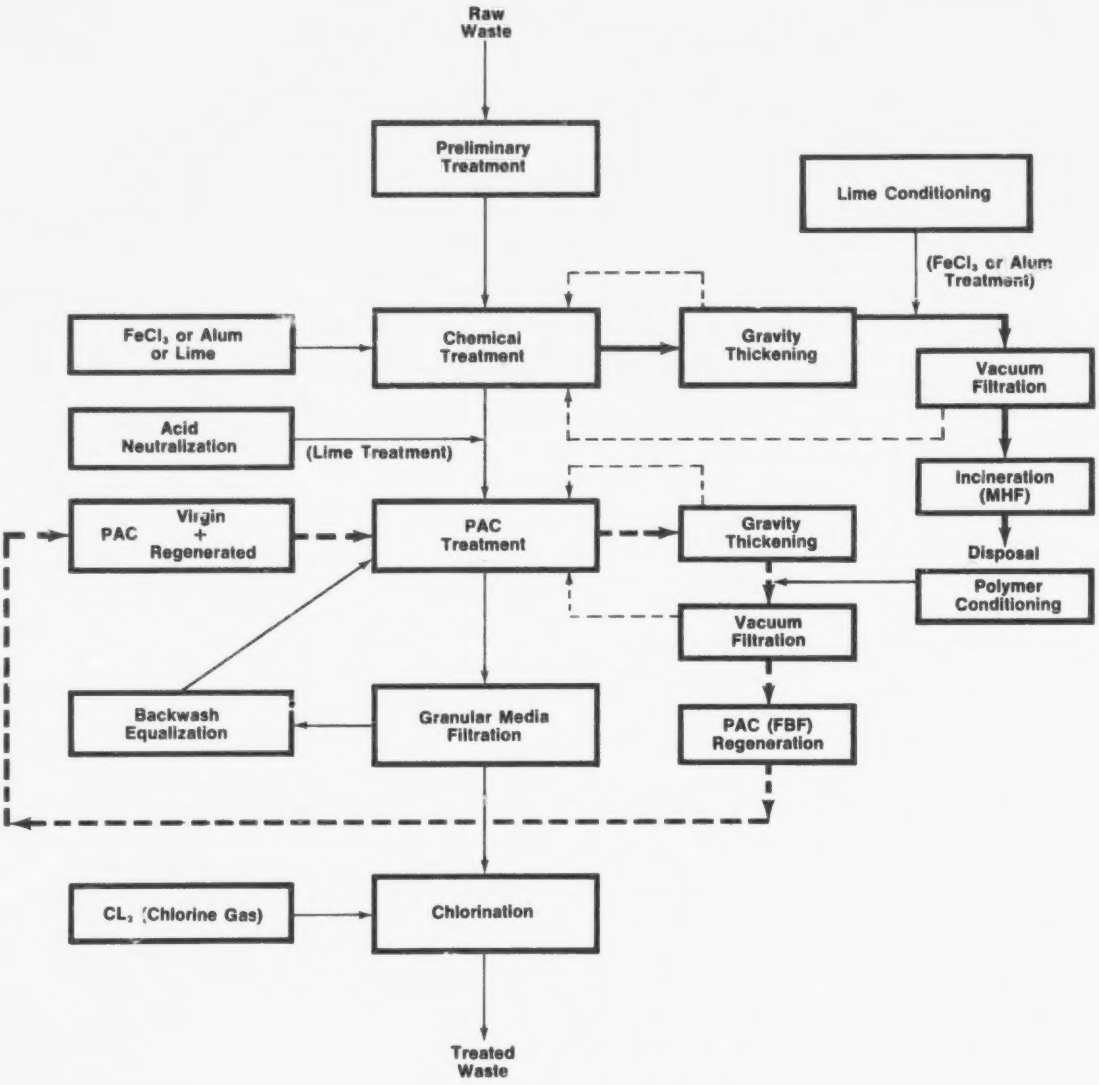
Raw Waste
Preliminary Treatment
Lime Conditioning
(FeCl$_3$ or Alum Treatment)
FeCl$_3$ or Alum or Lime
Chemical Treatment
Gravity Thickening
Vacuum Filtration
Acid Neutralization
(Lime Treatment)
Incineration (MHF)
PAC Virgin + Regenerated
PAC Treatment
Gravity Thickening
Disposal
Polymer Conditioning
Vacuum Filtration
Backwash Equalization
Granular Media Filtration
PAC (FBF) Regeneration
CL$_2$ (Chlorine Gas)
Chlorination
Treated Waste

TABLE 30: ASSUMED UNIT COSTS FOR ECONOMIC ANALYSIS

Chemicals:

Alum	3.5 ¢/lb
Ferric Chloride	5.0 ¢/lb
Lime ($Ca[OH]_2$)	1.0 ¢/lb
Acid (H_2SO_4)	1.0 ¢/lb
Polymer	40.0 ¢/lb
Powdered Carbon	10.0 ¢/lb
Chlorine	6.0 ¢/lb

Capital Costs:

All equipment costs were estimated as installed equipment by the authors firm and amortized at 7 percent interest for 20 years.

Operating Costs:

Power	3.0 $/MG
Sludge Incineration (16 hr /day):	
Alum & $FeCl_3$	3.4 $/ton of wet sludge
Lime	4.1 $/ton of wet sludge
Carbon Regeneration (16 hr /day)	3.4 $/ton of wet sludge
Ash Haulage	1.5 $/ton
Maintenance	1% of capital costs/year
Supervision and Operators[a] (seven men for the total PAC-PCT Plant)	87,600 $/year

---

[a] does not include incineration or regeneration operation costs

TABLE 31: MAJOR EQUIPMENT SIZES

CHEMICAL TREATMENT

	Alum	$FeCl_3$	Lime
Solids-Contact Units:			
Number	2	2	2
Diameter, ft	130	116	72
Gravity Sludge Thickener:			
Diameter, ft	56	44	44
Vacuum Filter:			
Number	3	2	2
Drum Size (diameter x face), ft	8 x 16	8 x 16	8 x 8
Sludge Incinerator (MHF)			
Diameter, ft	16'9"	16'9"	22'3"
No of Hearths	6	6	8

CARBON TREATMENT

Carbon Dosage, mg/ℓ	75	300
Solids-Contact Units:		
Number	4	4
Diameter, ft	90	90
Gravity Thickener		
Diameter, ft	20	40
Vacuum Filter:		
Number	1	1
Drum Size (diameter x face) ft	4 x 4	8 x 10
Regeneration Furnace		
Diameter, ft	10.75	18
Granular Media Filters		
Number	6	6
Diameter, ft	22	22

TABLE 32: ESTIMATED CHEMICAL TREATMENT COSTS (10 MGD PLANT)

Capital Costs: (1000 of $)	Alum	$FeCl_3$	Lime
Preliminary treatment	50	50	50
Chemical treatment	475	381	320
Gravity thickening	55	69	55
Vacuum dewatering	360	263	170
Sludge incineration	320	438	600
Total	1260	1201	1195
Contingency (50%)	630	600	598
Total capital cost	1890	1801	1793
Amortized cost, ¢/1000 gal.	4.9	4.7	4.6
Operating Costs: (¢/1000 gal.)			
Treatment chemical	4.1	5.0	3.8
Neutralization acid	0.0	0.0	2.8
Sludge conditioning chemical	0.2	0.3	0.0
Incineration O&M[a]	0.3	0.3	1.7
Power and maintenance	0.7	0.7	0.7
Supervision and labor	1.2	1.2	1.2
Total operating cost, ¢/1000 gal.	6.5	7.5	10.2
Total Treatment Cost, ¢/1000 gal.	11.4	12.2	14.8

a - O&M, operation and maintenance

about six times more dry sludge solids and about 3 times more water being fed to the incinerator for lime treatment than for alum or ferric chloride treatment.

It would be possible to reduce lime treatment neutralization costs by 1.5 to 2.0¢/1000 gal if the CO_2 content of incineration off-gas were used. Since incineration operation was scheduled for only 16 hr/day, stand-by acid or carbon dioxide feed facilities would be required. It is doubtful that lime-wastewater sludge recalcination could be justified for single-stage lime treatment of Salt Lake City wastewater. Limited $CaCO_3$ content of the sludge (about 60 percent of the dry solids) and increased furnace O&M costs would probably prevent economical recalcination and reuse of lime.

In summary, alum treatment of Salt Lake City raw municipal wastewater is the economic choice for chemical treatment. It should be noted that as wastewater feed phosphorus concentration increases and alkalinity decreases, the cost difference between alum and lime treatment would decrease.

POWDERED CARBON TREATMENT COSTS

Estimated powdered carbon treatment costs are preliminary due to the lack of definitive data on the PAC regeneration system performance. Table 33 shows preliminary estimated PAC costs. The regeneration O&M costs include operating labor and furnace utilities requirements. Power and maintenance and labor and supervision costs are estimated as one-half of that estimated for the complete PAC-PCT plant, excluding regeneration O&M costs. Pertinent process design parameters, based on results presented and discussed in Section VII, are as follows:

1. Mode of carbon contacting - two-stage counter-current
2. Carbon contactor average design overflow rate - 0.5 gpm/sq ft
3. Thickener solids loading - 20 lb/day-sq ft
4. Vacuum filter yield - 6.7 lb/sq ft-hr

TABLE 33: ESTIMATED POWDERED CARBON TREATMENT COSTS
(10MGD PLANT)

Carbon Dosage, mg/ℓ	75	300
CAPITAL COSTS: (1000 of $)		
Carbon Treatment	568	580
GM Filtration	228	228
Gravity Thickening	36	50
Vacuum Dewatering	25	72
Thermal Regeneration	225	450
Total	1082	1380
Contingency (50%)	541	690
Total Capital Cost	1623	2070
Amortized Cost, ¢/1000 gal.	4.2	5.4
OPERATING COSTS: (¢/1000 gal.)		
Carbon Make-up (15% losses)	0.9	3.8
Dewatering chemical	0.1	0.5
Regeneration O + M	0.1	0.4
Power and Maintenance	0.7	0.7
Supervision and Labor	1.2	1.2
Total Operating Cost, ¢/1000 gal.	3.0	6.6
TOTAL TREATMENT COST, ¢/1000 gal.	7.2	12.0

5. Filter cake moisture content - 78 percent

6. Carbon sludge conditioning polymer - 0.5 percent by wgt

7. Granular media filter average design rate - 3.0 gpm/sq ft

Table 33 shows costs for two dosages of PAC. These dosages were chosen to indicate the effect of dosage on total costs.

Organic removal is predicted using the two-stage countercurrent model in Table 23. The carbon system feed SCOD is assumed to be 50 mg/ℓ (recall that the chemical treatment step produced an effluent with 55-60 mg/ℓ of total COD). For the 75 mg/ℓ carbon dosage, C_o/M equals 50/75, or 0.67 mg/ℓ SCOD fed/mg/ℓ PAC and an X/M of 0.41 mg/ℓ SCOD removed/mg/ℓ PAC was computed. This results in a value of X equal to 0.41 x 75 or 31 mg/ℓ of SCOD removed and an effluent SCOD of 50-31, or 19 mg/ℓ. For the 300 mg/ℓ carbon dosage, C_o/M equals 50/300 or 0.167 mg/ℓ SCOD fed/mg/ℓ PAC and an X/M of 0.133 mg/ℓ SCOD removed/mg/ℓ PAC was computed. This results in a value of X equal to 0.133 x 300, or 40 mg/ℓ SCOD removed, and an effluent SCOD of 50-40, or 10 mg/ℓ. It is assumed that the granular media filter does not remove 5 mg/ℓ of particulate COD.

Predicted PAC-PCT plant effluent quality is less than 5 JTU turibidity, less than 5 mg/ℓ SS and 0.4 to 0.8 mg/ℓ phosphorus, for both carbon dosages. The plant effluent COD would be 24 and 15 mg/ℓ for 75 and 300 mg/ℓ carbon dosages respectively. Based on the BOD_5 data presented in Section VII, it was assumed that effluent BOD_5 would be about 12 and 8 mg/ℓ for 75 and 300 mg/ℓ carbon dosages respectively.

From the total treatment costs in Table 33 it is seen that a substantial increase in cost, 4.8¢/1000 gal. is required to produce an effluent COD of 15 mg/ℓ rather than 24 mg/ℓ.

The total treatment costs in Table 33 are based on assumed regeneration loss of 15 percent. Table 34 shows the economic impact of regeneration losses on total powdered carbon treatment costs. It is quite apparent that the higher the carbon dosage the greater the increase in total treatment costs as regeneration losses increase.

Table 35 shows the relative effect of assumed regeneration losses and carbon dosage on estimated powdered carbon regeneration costs. It is germain to note that even with up to 40 percent losses, it would be cheaper to regenerate and make-up losses with virgin carbon than to use virgin carbon only.

Using the capital and operating costs in Table 33, the total cost of powdered carbon treatment without regeneration was computed. For the 75 mg/ℓ carbon dosage, a total PAC treatment cost of 10.4 ¢/1000 gal. was determined. At 300 mg/ℓ carbon dosage, a total PAC treatment cost of 30.2 ¢/1000 gal. was determined. As plant size (mgd) decreases below 5 mgd, the unit capital and operating costs increase substantially while virgin carbon costs remain essentially constant. A rigorus economic analysis would identify a given plant size below which powdered carbon regeneration would not be economically feasible, especially at a nominal carbon dosage of about 100 mg/ℓ. A rough estimate of this plant size would be in the 1 to 2 mgd range.

TOTAL PAC-PCT TREATMENT COSTS

An indication of total PAC-PCT process treatment costs for Salt Lake City municipal wastewater is obtained from data presented in Tables 32 and 33. An additional cost of 0.4 ¢/1000 gal. was estimated for chlorination of the plant effluent with 5 mg/ℓ of chlorine.

Alum treatment followed by two-stage counter-current carbon contacting with 75 mg/ℓ dosage would cost 18.3 ¢/1000 gal. The predicted plant effluent quality would be considerably better than a secondary biological treatment effluent for all parameters, but

TABLE 34 ESTIMATED POWDERED CARBON TREATMENT COSTS FOR VARIOUS REGENERATION LOSSES

Carbon Dosage, mg/ℓ	75	300
Assumed Regeneration Losses, %	Total Treatment Costs, ¢/1000 gal.	
0	6.3	8.2
5	6.6	9.5
10	6.9	10.8
20	7.5	13.4
30	8.1	16.0
40	8.7	18.6
50	9.3	21.2

TABLE 35: ESTIMATED POWDERED CARBON REGENERATION COSTS

Carbon Dosage, mg/ℓ	75	300
Assumed Regeneration Losses, %	Total Regeneration Costs ¢/lb of PAC used	
0	2.1	1.2
5	2.6	1.8
10	3.0	2.3
20	4.0	3.3
30	5.0	4.4
40	5.9	5.4

especially with respect to phosphorus and suspended solids. It is doubtful whether secondary biological treatment followed by tertiary treatment for phosphorus removal could be accomplished for less than 18.3 ¢/1000 gal.

SECTION IX

ADDITIONAL STUDIES

The results of the present study have demonstrated the operability and treatment effectiveness of the PAC-PCT process. Estimated economics indicate that the process may be an economic alternative to presently available treatment approaches for certain wastewaters. However, several areas need further study to develop new and/or more reliable design data. Because of the above factors, a follow-on contract was granted for one additional year of studies.

The major objectives of these studies are as follows:

1. Determine the effect of diurnal flow variations on plant performance, especially the solids-contact treatment units.
2. Attempt to quantify the effect of biological activity in the carbon contactors on organic removal.
3. Continue development of the fluidized bed furnace to minimize carbon losses.
4. Determine the properties of regenerated carbon, especially its effectiveness in the carbon treatment system.
5. Obtain pilot plant scale results for gravity thickening and vacuum filter dewatering of chemical-sewage sludges and compare same with laboratory test results.
6. Evaluate different granular media filter bed designs to improve the efficiency of filter operation; also evaluate the effect of various backwash parameters.
7. Evaluate the use of continuous on-line ultra violet adsorption analysis of carbon system feed and effluent to determine if carbon can be fed at a minimum rate necessary to produce a uniform quality feed effluent.

8. Obtain long term plant performance and operating data.

SECTION X

ACKNOWLEDGEMENTS

The support and indulgence of the Salt Lake City Commissioner of Water and Sewers, city engineers and municipal sewage pumping plant personnel is acknowledged.

Special acknowledgement is due Messrs. Jim Snarr, Darrell Cook and Richard Wallace who nursed the pilot plant through start-up and kept it operating under adverse conditions.

The support, counsel and patience of the project officer, Mr. James Westrick and Mr. Jesse M. Cohen of the Advanced Waste Treatment Research Laboratory, National Environmental Research Center Cincinnati, Ohio, Environmental Protection Agency is acknowledged with sincere thanks.

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SECTION XIII

ABBREVIATIONS AND SYMBOLS

Abbreviations:

BOD_5	five-day, 20°C-Biochemical Oxygen Demand
COD	Chemical Oxygen Demand
CT	Vacuum filter drum cycle time
FT	Vacuum filter cake form time
ft	feet
gal	gallon
g	gram
GMF	granular media filter
gpd	gallons per day
gpm	gallons per minute
hr	hour
JTU	Jackson turbidity units
Kg	kilogram
ℓ	liter
lb	pound
Log	logarithm
mgd	million gallons per day
mg	milligram
MG	million gallons

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MHF	multiple hearth furnace
min	minutes
ml	milliliter
mm	millimeter
O&M	operation and maintenance
ΔP	incremental pressure
PAC	powdered activated carbon
PCT	Physical-Chemical Treatment
P/I	pressure indicator
rpm	revoluations per minute
SCOD	soluble chemical oxygen demand
SCFM	standard cubic feet per minute (@ 20°C, sea level)
sec	second
SG	specific gravity
sq	square
SRT	solids retention time
SS	suspended solids
STOC	soluble total organic carbon
TOC	total organic carbon
Temp	temperature
W	vacuum filter dry cake weight
wgt	weight
2SCC	two-stage counter-current
K_1, K_2	constants in Equation (6)
A, n	constants in Equation (7)

Symbols:

@	at
°C	degrees Centigrade
C_o	feed organic concentration
C_i	intermediate stage organic concentration
C_e	effluent organic concentration
°F	degrees Fahrenheit
M	powered carbon concentration
X	organics removed (C_o-C_e)
Θ_D	vacuum filter cake dry time
¢	cents
$	dollars
∞	infinite
∝	proportional to

SECTION XIV
APPENDICES

APPENDIX A

ADSORPTION OF SOLUBLE WASTEWATER ORGANICS ON ACTIVATED CARBON

Adsorption is defined as the concentration of material at an interface between two phases, one of which has the capability of dispersing the material. Applied to soluble wastewater organic material and activated carbon, adsorption is the concentration of soluble organics at the water-carbon interface or the activated carbon surface. Based on the chemical and physical properties of the organics, water, and carbon surface, there is a unique distribution of organics between the bulk of the water and the water-carbon interface.

There are several theoretical models which have been used to describe the adsorption phenonmenon. None have been shown universally applicable to wastewater-activated carbon systems. This is not surprising, considering the extremely heterogenous nature of soluble organics in wastewaters (molecular size, concentration, functionality, etc.) and of activated carbon (pore size and distribution, surface functionality).

The emperical Freundlich model has been widely used for wastewater-activated carbon adsorption systems[14,15,23,34,35]. However, its complete adequacy has not been fully established. Two major shortcomings are that a maximum adsorption capacity (loading) or any unadsorbable soluble organic fraction are not accounted for.

When evaluating relative effects of carbon type, pretreatment and method of contacting, the above shortcomings do not limit utility of the Freundlich model.

In as much as a laboratory and pilot plant soluble organic removal data were evaluated by the Freundlich model and that multi-stage counter-current carbon contacting was evaluated, the following is a presentation of the model and the impact of counter-current contacting.

Figure A-1 is a log-log plot of SCOD adsorbed from Salt Lake City raw wastewater (after 0.45 micron membrane filtration) onto Aqua Nuchar A carbon versus equilibrium SCOD concentration. These data were developed from a laboratory equilibrium isotherm test.

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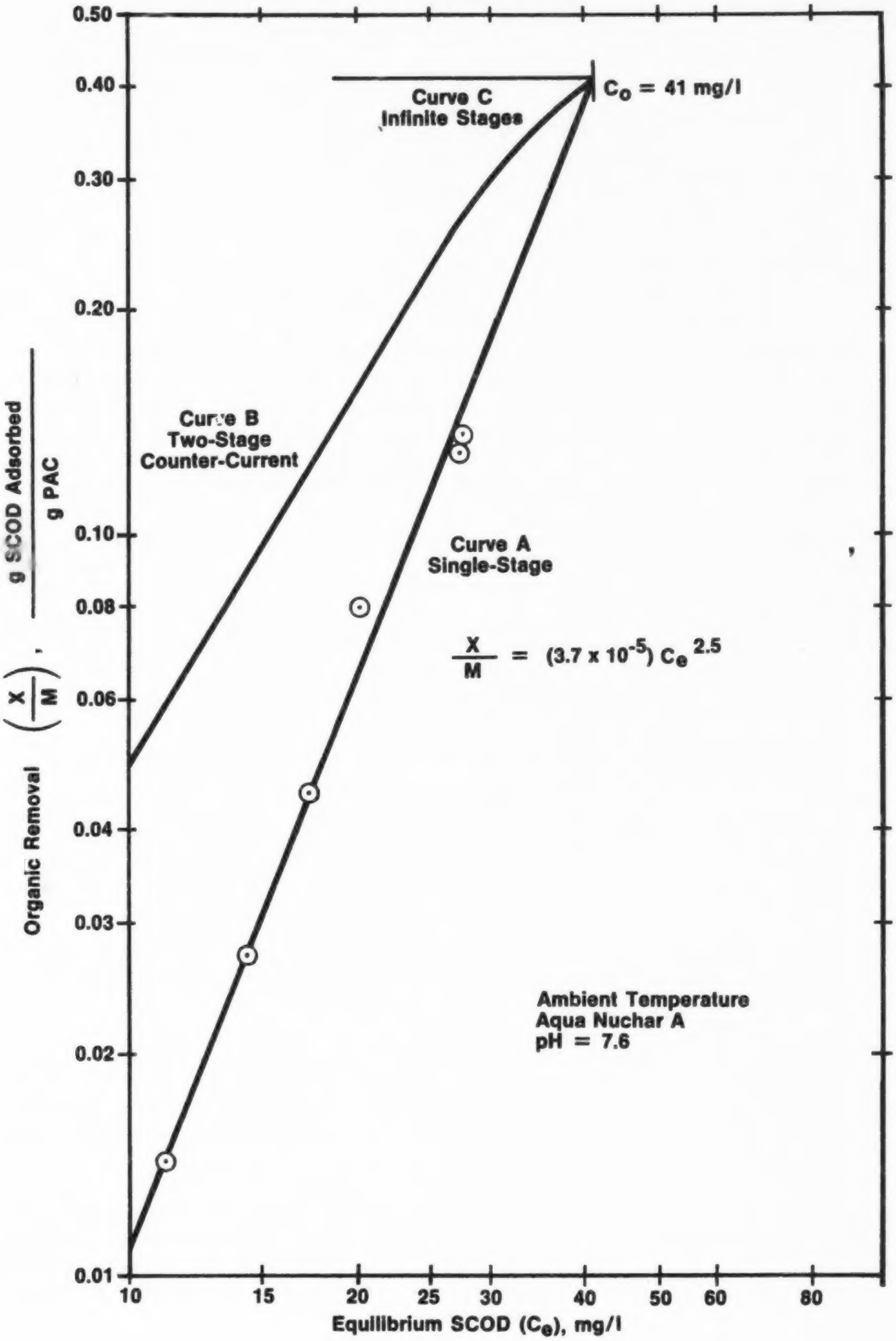
FIGURE A-1 : ADSORPTION OF SOLUBLE COD ON CARBON



The data can be described by the Freundlich model, (Curve A):

$$X/M = A\ C_e^n \qquad (7)$$

Where: X = Weight of organics adsorbed
M = Weight of carbon
C_e = Equilibrium organic concentration in the bulk solution
A, n = Are empirical constants

For the data shown in Figure 60, the constants A and n are 0.000037 and 2.5 respectively, for C_e in mg/ℓ of SCOD and X and M in similar weight units. If it is assumed that there is no "unadsorbable fraction" of SCOD and that the maximum possible adsorption capacity of the carbon used is greater than 0.40 lbs SCOD/lb PAC (i.e., X/M at C_o for Curve A), then the equation,

$$X/M = 0.000037\ (C_e)^{2.5} \qquad (8)$$

will describe a single-stage adsorption system as depicted in Figure A-2.

From Equation (8) the concentration of PAC required to produce any desired system effluent C_e can be computed. For example, if a 15 mg/ℓ SCOD effluent (C_e) is required and C_o is 41 mg/ℓ SCOD, then from Equation (8) X = 41-15 = 26 mg/ℓ SCOD and M is found to be 26/0.031 = 850 mg/ℓ PAC. This high carbon dosage requirement is due to the low loading, X/M, at a C_e of 15 mg/ℓ . A significantly reduced carbon dosage could be used if a treatment scheme were used where the carbon was equilibrated at a higher C_e value.

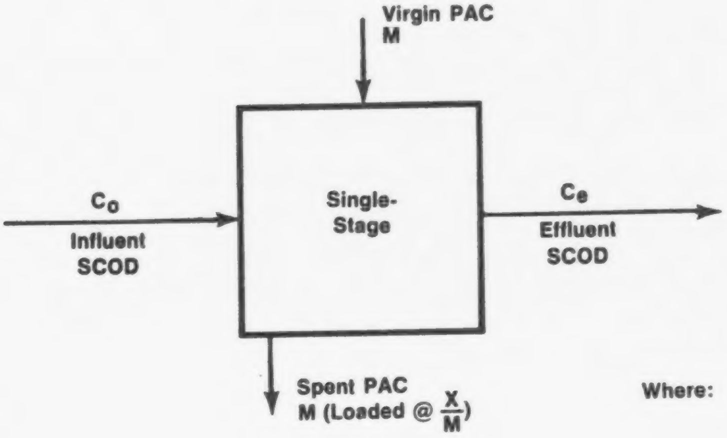
The two-stage counter-current treatment scheme depicted in Figure A-3 will reduce the concentration of carbon required to produce a given effluent Ce, compared with the single-stage system. This results from the carbon leaving the adsorption system being equilibrated at C_i which is greater than C_e.

If it is assumed that the chemical and physical characteristics of the organics (SCOD) are identical throughout the adsorption system, the following development will lead to a model of the two-stage counter-current system:

Given:

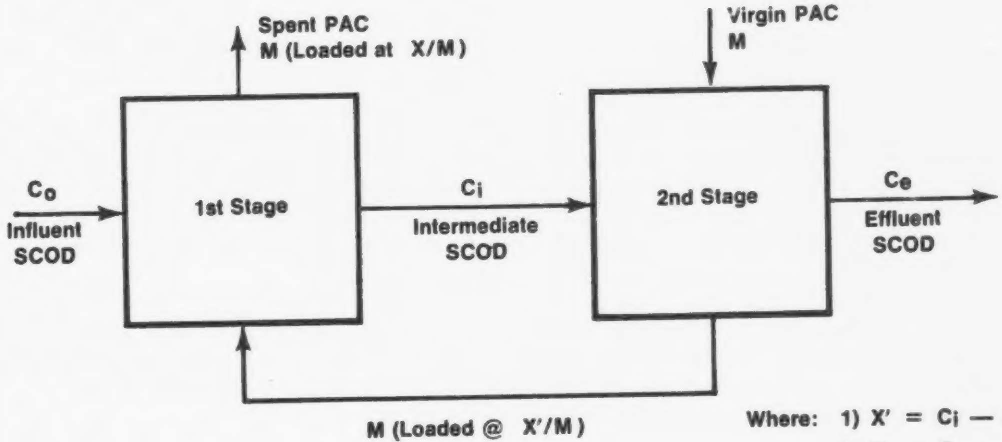
$$X/M = A\ (C_i)^n \qquad (9)$$
$$X'/M = A\ (C_e)^n \qquad (10)$$
$$X' = C_i - C_e \qquad (11)$$
$$X = C_o - C_e \qquad (12)$$

FIGURE A-2 : SINGLE-STAGE CARBON ADSORPTION SYSTEM


Virgin PAC
M
C_o
Influent
SCOD
Single-
Stage
C_e
Effluent
SCOD
Spent PAC
M (Loaded @ $\frac{X}{M}$)

Where: 1) $X = C_o - C_e$
2) all units, mg/l

FIGURE A-3 : TWO-STAGE COUNTER-CURRENT CARBON ADSORPTION SYSTEM


Spent PAC
M (Loaded at X/M)
Virgin PAC
M
C_o
Influent
SCOD
1st Stage
C_i
Intermediate
SCOD
2nd Stage
C_e
Effluent
SCOD
M (Loaded @ X'/M)

Where: 1) $X' = C_i - C_e$
2) $X = C_o - C_e$
3) all units, mg/l

Solving Equation (10) for A and substitution of its value in Equation (9) will result in:

$$X/M = \frac{X'}{M}\frac{(C_i)^n}{(C_e)^n} \qquad (13)$$

Substitution of the values for X and X' from Equations (11) and (12) into Equation (13) will give:

$$C_o - C_e = \frac{(C_i - C_e)(C_i)^n}{(C_e)^n}$$

rearranging, $(C_i)^{n+1} - (C_e)(C_i)^n = (C_o)(C_e)^n - (C_e)^{n+1}$ (14)

For a given wastewater - carbon system (i.e., C_o, A and n), it is seen that C_i is a non-linear function of C_e. Therefore, the two-stage counter-current system cannot be modeled by a Freundlich equation. In other words, a log-log plot of X/M versus C_e for the two-stage response will not be a straight line in Figure A-1.

Using Equations (9) through (12) and A, n and C_o from Figure A-1, a trial and error solution to Equation 14 was used to determine the X/M versus C_e relationship for a two-stage counter-current system. This relationship is shown as Curve B in Figure A-1. The same assumptions made for Equation (8) hold for this relationship.

Using a two-stage counter-current adsorption system to produce an effluent of 15 mg/ℓ SCOD (for C_o = 41 mg/ℓ SCOD) would require 260 mg/ℓ of PAC as opposed to 850 mg/ℓ when using a single-stage adsorption system.

For an infinite number of counter-current contact stages, the carbon loading would be constant for any value of C_e (Curve C shown in Figure A-1). The PAC would be equilibrated with the highest possible C_e value (i.e., C_o). It is theoretically possible to approach infinite stage efficiency by employing packed-bed adsorption columns[36]. For such a system the adsorption relationship shown in Figure A-1 would indicate a carbon loading of 0.40 lb SCOD/lb PAC. In summary, the following carbon dosages would be required for producing a 15 mg/ℓ SCOD effluent.

# of Counter-Current Contacting Stages	PAC Dosage, mg/ℓ
1	850
2	260
Infinite	66

Some investigators have used laboratory equilibrium adsorption isotherms to predict organic removals for granular carbon column applications. The normal procedure used is to extrapolate the equilibrium test results to the feed organic strength (C_o) and observe the organic removal (X/M) at that point. That this procedure results in imprecise conclusions, even when precise equilibrium adsorption isotherm data is used, can be shown by the following example. The equilibrium data for Curve A in Figure A-1 were statistically analyzed. A linear regression of log X/M on log C_e resulted in a 0.99 correlation coefficient, indicating a very precise fit of the data. Extrapolation of the regression curve 13 SCOD units, to a C_o of 41 mg/ℓ, indicated a X/M loading of 0.40 g SCOD/g PAC. However, the 95 percent confidence interval for this "estimated" value of X/M at C_o is from 0.28 to 0.55 g SCOD/g PAC, indicating that extrapolation of very precise equilibrium data can result in imprecise estimates of organic loading.

APPENDIX B

WATER QUALITY PARAMETERS AND TEST PROCEDURES

TURBIDITY

Turbidity of grab and composite samples were determined with a Hach Chemical Company, Model 2100 laboratory turbidimeter. This device is a nephelometer (light scattering) which was standardized with a plastic rod. Turbidity is reported as Jackson Turbidity Units (JTU).

Caution must be exercised when interpreting turbidity of samples containing powdered carbon. Since powdered carbon absorbs white light, the normal relationship between turbidity and suspended solids does not exist. Figure B-1 shows that powdered carbon concentration is directly related to the amount of transmitted light but not to turbidity.

The plant process turbidimeter used as a Keene Instruments Company flow-through device which measured both scattered and transmitted light. The process turbidimeter was routinely calibrated against the laboratory unit to facilitate comparison of results.

SUSPENDED SOLIDS

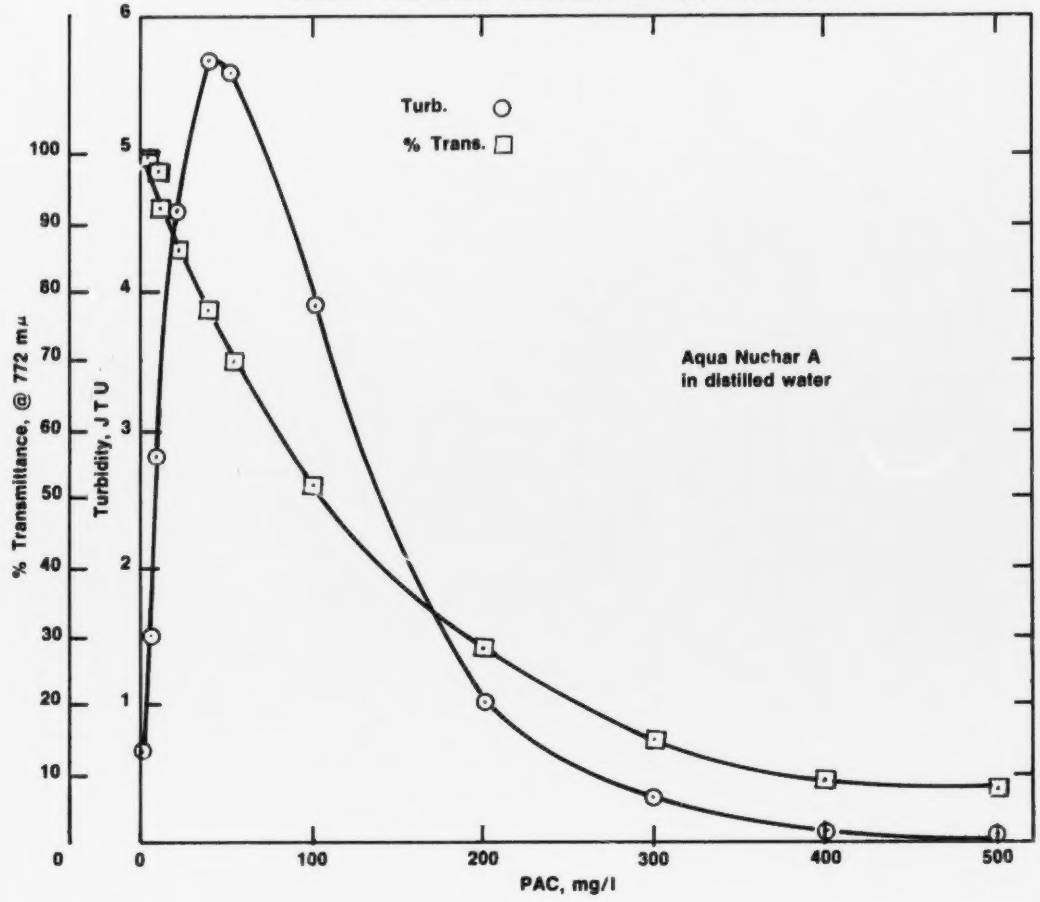
Suspended solids were defined as material retained by a 0.45 micron membrane filter. The Standard Methods[37] procedure for Nonfilterable Residue was used. It should be noted that for the sample sizes used, samples with less than 10 mg/ℓ SS had weighing errors greater than about 10 percent.

Concentrated samples of chemical-sewage sludge and PAC slurries were filtered through glass mat filter pads (Type GFB).

PHOSPHORUS

Total Phosphorus was determined by the persulfate method[37]. Soluble total phosphorus was defined as that phosphorus passing through a 0.45 micron membrane filter.

FIGURE B-1 : POWDERED CARBON:
TURBIDITY AND LIGHT TRANSMITTANCE OF SUSPENSIONS



TOTAL ORGANIC CARBON

Total and soluble organic carbon (TOC, STOC) was determined with a Beckman Instruments Company Total Organic Carbonaceous Analyzer. Prior to analysis, inorganic carbonates were removed from the samples by acidification and stripping with inert gas.

CHEMICAL AND BIOCHEMICAL OXYGEN DEMAND

Total and soluble chemical oxygen demand (COD, SCOD) and 5-day, 20°C BOD_5 were determined according to Standard Methods[37]. Total COD was not determined on carbon contactor effluent samples since one mg/ℓ of powdered carbon was found to exert about 1 mg/ℓ COD.

pH

The pH data reported were obtained with a digital readout laboratory pH meter. The process pH meter and recorder were used only to automatically control acid neutralization of lime treatment effluent. The plant instrument was calibrated against the laboratory unit which was calibrated with standard buffered solution.

HARDNESS, ALKALINITY, SULFIDES, IRON, ALUMINUM, AND AMMONIA NITROGEN

These parameters were determined according to Standard Methods[37].

AVAILABLE LIME INDEX

"The available lime index" of hydrated lime was determined using the procedure presented in "Standard Analysis of Limestone, Quicklime, and Hydrated Lime", ASTM, Part, 9, 1964 C24-58, p.22.

OXYGEN

Oxygen content of furnace off-gases was analyzed by means of a Model 715 Process Oxygen Monitor manufactured by Beckman Instruments, Inc., Fullerton, Cailfornia.

TOTAL SUSPENDED SOLIDS, ASH, AND VOLATILE SOLIDS

Total suspended solids of carbon slurry samples was determined according to Standard Methods[37]. Ash content of carbon solids was determined according to the procedure presented in ASTM, Part 28 (1964), D1506-59, p. 750. Volatile solids of carbon samples was determined according the the procedure presented in ASTM, Part 28 (1964), D1620-60, p. 811. Numerous volatile solids determinations of spent and regenerated carbon samples were invalid due to burning of some fixed carbon. The problem appeared to be caused by several factors. The depth of carbon cake placed in the oven, of up to 1 inch, didn't allow good heat distribution. Also, the time the sample was left in the furnace was critical. Too long a time resulted in burning of carbon. These problems were recognized late in the study.

JAR TEST PROCEDURE

A six place Turbitrol Jar Test Apparatus, made by the Taulman Company was employed. Sample volumes of 1.0 or 1.5 liters were used. The following procedure was established as a "standard":

1. Measure the wastewater sample volume in a 2-liter graduate cylinder and pour into the jar test beaker.

2. Start rapid mixing of samples at 100 rpm and add desired chemical (aluminum, ferric salt or lime) and continue mixing for one (1) minute. [Note! If a polymeric flocculation aid was used, it was added just prior to the end of the 100 rpm mixing period.]

3. Reduce mixing speed to 30-50 rpm, making sure that the majority of floc particles are maintained in suspension, and slow mix for 10 minutes.

4. After slow mixing, the samples were allowed to quiescently settle for 10 minutes. If floc settling rate was to be determined, the height of the solids (solids-liquid interface) was noted at appropriate time intervals, or the time at which the majority of floc had settled to the bottom of the beaker was noted. In the former case, a plot of height versus time was made and floc settling rate, in inches/minutes, obtained from the linear portion of the curve after the influence of initial mixing disturbance was eliminated.

5. After the 10 minutes of quiescent settling period, a supernatant sample was withdrawn one (1) inch below the liquid surface for desired analytical testing (e.g., turbidity, pH, total and/or soluble phosphorus, total and/or soluble COD, hardness, etc.).

6. When floc settling rates were to be determined, the settled floc were resuspended, after Step 5, by slow mixing at 30-50 rpm for 2-3 minutes and the settling rate of floc noted again. The average settling rates of the first (Step 4) and second (Step 6) test was reported.

In an attempt to simulate solids-contact treatment, an alteration of the above procedure was used:

7. After Step 5 or 6, decant as much clear liquor as possible, leaving the concentrated settled solids in the beaker.

8. With the sludge in Step 7, repeat Steps 1 through 7. In addition to any or all of the data previously noted, measure the depth of sludge present after each contact.

9. Repeat Step 8 for 4 to 8 times.

A graphical plot of floc settling rate, supernatant turbidity or other parameters are plotted versus number of contacts (e.g., number of times Step 8 is repeated plus one) to indicate any effect of simulated solids-contact treatment.

During Phase I of this study, laboratory reagent grade chemicals were used. After pilot plant start-up, jar tests were usually conducted using plant chemical supplies.

EQUILIBRIUM ADSORPTION ISOTHERMS

Suitably sized raw wastewater samples were chemically pretreated as desired and the following procedure used, employing the six place Turbitrol Jar Test Apparatus:

1. Acid rinse all glassware.

2. Prepare powdered carbon samples as follows:

   a. tare five 100 ml beakers.
   b. add a desired weight of oven dryed (110°C) carbon to each of the beakers and add 20-30 ml of distilled water.
   c. place beakers, covered with watch glasses, on a hot plate and bring contents to a boil, then remove from hot plate and cool to near ambient temperature.

3. Add 1.0 to 1.5 liter samples of chemically pretreated and 0.45 micron membrane filtered wastewater to each of six beakers and mix at 100 rpm.

4. Measure the initial temperature and pH of each sample, adjusting the pH with concentrated H_2SO_4 or NaOH if desired (normal test pH was 7.0-7.5).

5. To the mixing samples, effect a quantitative transfer of degassed powdered carbon (from Step 2), recording the amount of distilled water used. A blank wastewater sample was always run without any carbon.

6. Mix samples for one (1) hour at 100 rpm (preliminary tests had indicated that at least 95 percent SCOD removals was obtained at 15-20 minutes contact time).

7. At the end of one (1) hour determine pH and temperature of each sample.

8. Filter a suitable sized sample from each beaker through a 0.45 micron membrane filter and determine the organic concentration (as measured by COD or TOC analysis).

Standard procedure involved running triplicate organic concentration tests on the blank and duplicate tests on each carbon contact sample. The difference between the blank and carbon contact organic concentration was considered adsorbed organics. The organic removal (mg/ℓ adsorbed organics/mg/ℓ carbon) was plotted versus the equilibrium organic concentration (mg/ℓ) on log-log graph paper.

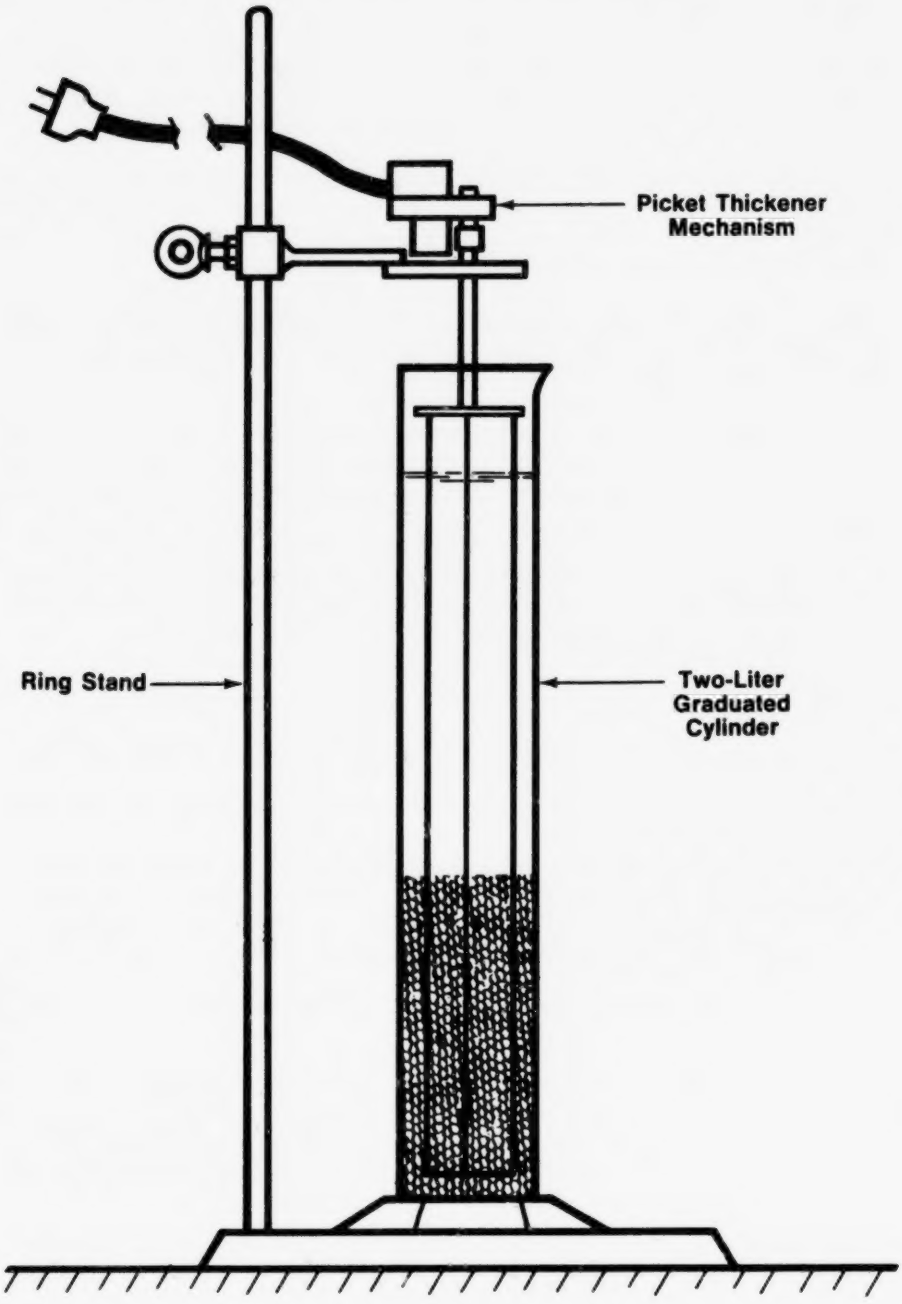
GRAVITY THICKENING TESTS

The gravity thickening characteristics of sludge samples were determined by the following test procedure using the laboratory test device shown in Figure B-2:

1. A 24-hour composite of sludge blowdown was obtained. If tests were to be run at different initial solids concentrations, then a sample of clarifier overflow was also obtained for dilution of the sludge.

2. Pour two liters of representative sludge into the 2-liter graduated cylinder for which the volume versus depth relationship is known.

FIGURE B-2 : SLUDGE THICKENING TEST APPARATUS


Picket Thickener Mechanism
Ring Stand
Two-Liter Graduated Cylinder

3. Gently mix the contents to insure uniform suspension (a rubber stopper on the end of a rod works well). If a conditioning chemical was used, it was added during this gentle mixing.

4. Immediately insert the picket thickener mechanism into the sludge and start it rotating at 6 revolutions per hour. Also, immediately start a clock timer.

5. Observe and record sludge interface height (milliliters) at appropriate time intervals such that a smooth curve is identified. The test is normally run until no further thickening is observed (8-24 hours).

6. Remove the picket mechanism and decant as much clear supernatant as possible recording the volume of thickened sludge.

7. Filter all of the thickened sludge through a Whatman #1 filter paper, using a Buchner funnel, saving the filtrate for a determination of supernatant specific gravity.

8. Dry the filtered sludge at 110°C to constant weight to determine the mass of sludge solids (supernatant from Step 6 is assumed to contain no SS).

9. Determine thickener loading by the following method:

   a. Plot sludge interface height versus time as in Figure B.3.
   b. Draw tangents AB and CD and bisect their obtuse angle as per line EF.
   c. Construct line GH perpendicular to line EF and tangent to the thickening curve. Curve AH is the "working line" which is used to compute predicted solids loading at a given underflow concentration.
   d. Compute predicted full scale thickener loading (U):

$$U = \frac{C_o H_o K}{t_x} \qquad (15)$$

where: C_o = initial suspended solids concentration, lb (dry solids)/cu ft

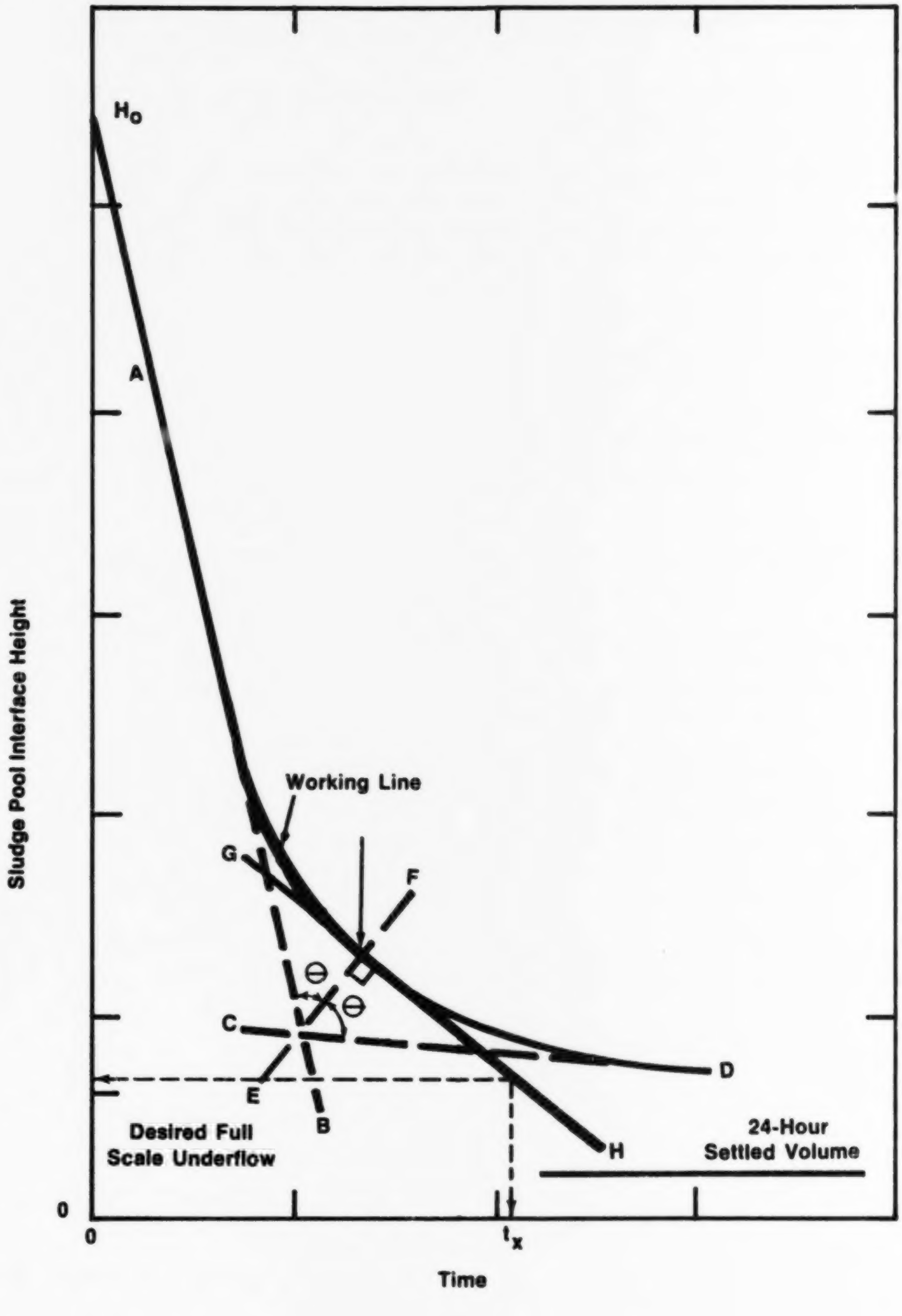
H_o = initial sample depth, ft

t_x = from Figure B.3, days

K = desired scale up factor, normally about 0.8

U = is reported in lb (dry solids) per day per square foot of thickener area.

FIGURE B-3: TYPICAL SLUDGE THICKENING CURVE



H_o
A
Working Line
G
F
Θ
Θ
C
D
E
B
H
Desired Full Scale Underflow
24-Hour Settled Volume
Sludge Pool Interface Height
0
0
t_x
Time

The laboratory thickening data was reduced in the following manner:

1. A desired underflow concentration in percent solids by weight was assumed.

2. Knowing the specific gravity of the solids (SG_s) and supernatant (SG_ℓ) and the weight of solids (W_s), the volume of slurry ($V_{s\ell}$) associated with the assumed underflow concentration (%S) was computed as follows.

$$V_{s\ell} = \frac{W_s}{SG_s} + \frac{W_\ell}{SG_\ell} \qquad (16)$$

Where: W_ℓ = weight of liquid in underflow

$$= \frac{W_s(100)}{\%S} - W_s \qquad (17)$$

3. The sludge pool interface height associated with the computed volume of slurry was determined.

4. The value of t_x associated with volume of slurry via the working line (AH) in Figure B-3 was observed and the predicted full scale thickener solids loading (U) computed using Equation (15).

5. Repeat Steps 1 through 4 for several values of assumed underflow concentrations, thus generating data to provide a smooth curve of thickener solids loading vs underflow solids concentration for a given sample of sludge (i.e., initial solids concentration).

Laboratory thickening test results shown in Figures 23, 26 and 33 were developed in the above manner.

VACUUM FILTER LEAF TESTS

Leaf tests were conducted on thickened 24-hour composite sludge samples. The filter leaf used had an area of 0.10 sq ft. The following types of filter media were used for the various sludges tested:

Sludge	Media
Lime-Sewage	NY-527-F, POPR-853, POPR-852-F, POPR-853-F
$FeCl_3$-Sewage	PO-801-HF, PO-802-HF
Alum-Sewage	NY-527-F
Spent Carbon	NY-432-F, NY-527-F, POPR-873

The general arrangement of laboratory equipment for conducting the leaf test is shown in Figure B-4. The procedure followed was:

1. Place about 1.5 liters of a representative sample of sludge in a 2 liter beaker.

2. If conditioning chemicals are used, they are gently mixed into the sludge with a large spatula.

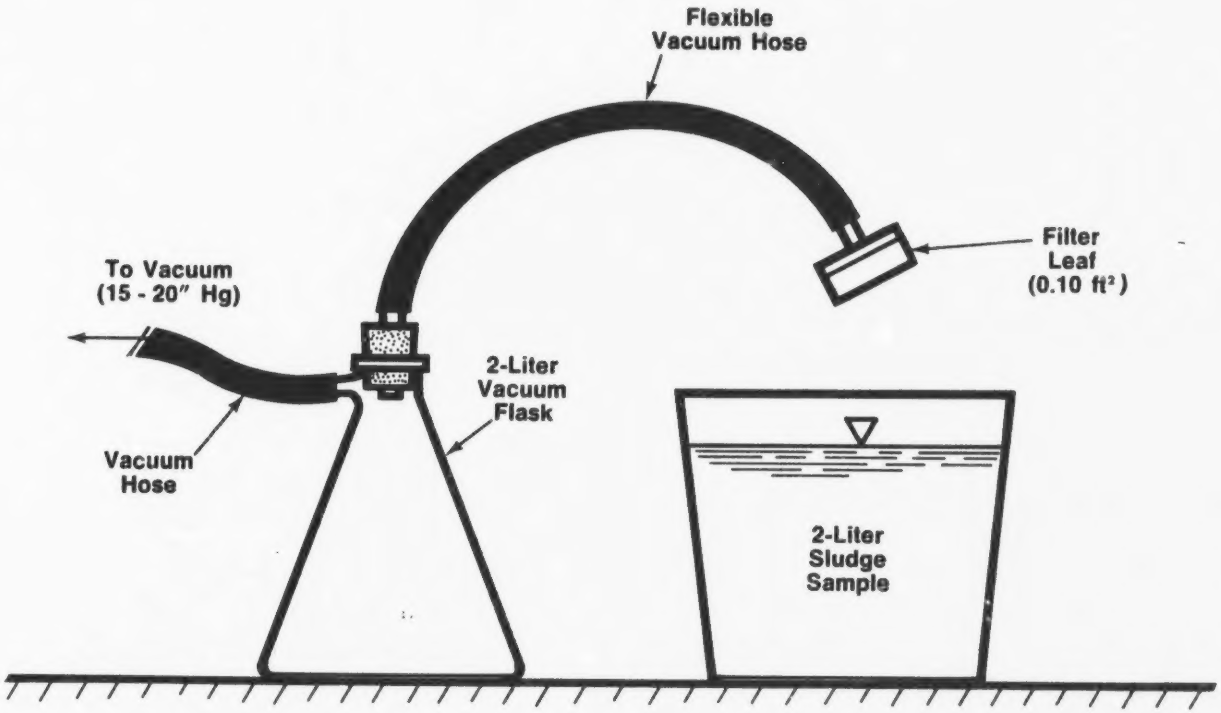
3. Apply a vacuum to the filter leaf, normally 15-20 inches of Hg.

4. Immerse the properly sealed and conditioned filter leaf into the mixed sludge for a desired cake form time (FT) normally 1/2 to 4 minutes.

5. Remove the leaf from the sludge and maintain the vacuum for a desired cake dry time (Θ_D) normally 4 to 1/2 minutes. (Note! Dry time used is inversely proportional to form time).

   a. If cake cracks, note the time.
   b. Note any reduction in vacuum level during dry time.

FIGURE B-4 : SLUDGE DEWATERING TEST APPARATUS


Flexible Vacuum Hose
Filter Leaf (0.10 ft²)
To Vacuum (15 - 20" Hg)
2-Liter Vacuum Flask
Vacuum Hose
2-Liter Sludge Sample

6. Turn off vacuum at end of dry time.
7. Note any difficulty in physically removing cake from leaf.
8. Place all of cake on a tared evaporation dish and weigh immediately. This will give wet filter cake weight.
9. Measure filter cake thickness.
10. Dry the filter cake to constant weight at 110°C - this will give the dry filter cake weight (W).
11. Determine filtrate volume (contents of vacuum flask) and filtrate suspended solids.
12. Repeat Steps 3 through 11 for various (usually 4 or 5) combinations of form and dry times.

From the above data the maximum full scale filter yield (lb dry solids/sq ft-hr) was determined for producing a dischargable cake thickness, assuming a desired drum submergence and scale up factor (normally 33 percent submergence and 0.8 scale up factor).

APPENDIX C

OPERATION OF SOLIDS-CONTACT UNITS

Operation of solids-contact treatment units used in this study involved monitoring and controlling the inventory of solids within the unit. There were four zones within the unit where control was desirable (letters refer to Figure 15):

1. Initial mixing zone or draft tube (E).
2. Secondary mixing zone or reaction zone (Q).
3. Clarification zone (J).
4. Sludge removal zone (I).

Solids monitoring was routinely conducted at all zones except the initial mixing zone. Since the contents of the initial mizing zone flows into the secondary mixing zone, the solids content of the two zones were considered to be equivalent. Solids concentrations were determined periodically by conducting suspended solids tests on samples from the reaction zone, the three clarification zone sample taps and sluge blowdown. Because of the time and expense required to make suspended solids determinations, an indirect measure of solids concentration was employed as a routine operational tool. This method, referred to as a 5-minute settling test, consisted of obtaining a two(2) liter sample in a 2-liter graduated cylinder. The sample was allowed to settle quiescently for exactly five (5) minutes and the height of settled sludge observed. The percent volume of sludge settled in five minutes was found to be precisely related to the slurry concentration.

Two aspects of solids inventory were considered important. First, the concentration of solids in the different zones was important. Generally speaking, a relatively high solids concentration was desired in all zones but the clarification zone. Second, the total solids mass within the unit was important. For a given solids inventory (mass), the solids flux through the treatment unit established the mean solids residence time (SRT). The solids flux depended upon flow, feed solids removed and chemical precipitates removed. In wastewater treatment applications, retention of biodegradable material for excessively long times may result in anaerobic biological activity with possible odor and gassification. Therefore, control of SRT by monitoring and adjustment of solids inventory was desirable. The concentration of solids and the volume of slurry establish the solids inventory within a treatment unit. For a given volume of slurry, the higher the solids concentration, the larger the sludge inventory. The volume of sludge in the solids-contact units was approximated by withdrawing samples from the clarification zone sample taps and noting the highest tap at which sludge was present.

Operation of the solids-contact unit to control solids concentration and inventory involved manipulation of the pumping turbine and sludge rake rotational speeds, and the sludge blowdown frequency and duration. Increasing pumping turbine speed increased the sludge pumpage, or circulation rate and usually resulted in a more concentrated suspension being circulated. Normally, a given pumpage rate exists above which no increase in solids concentration or benefit of additional contacting is obtained. In addition, there is a maximum turbine tip speed desirable depending on the type of floc solids being handled. High tip speeds may shear previously formed floc particles.

The desired sludge rake rotational speed depends upon the rate of solids circulation and blowdown. Generally speaking, the higher the pumping turbine speed, the higher the rake speed required. In addition, the higher the solids flux through the treatment unit, the higher the required rake speed to move settled solids to the sludge outlet. An interacting factor is the extent to which settled sludge thickens. Obviously, the more concentrated the sludge, the lower the rake speed required to transfer a given quantity of sludge from the clarification zone to the draft tube inlet and sludge outlet. It should also be noted that an excessive rake speed will hinder sludge concentrating.

Sludge blowdown was accomplished automatically by an adjustable timer controlled blowdown valve. An available static head of about 16 feet of water provided the driving force and the automatic control simply activated the sludge valve air operator. In order to maximize sludge blowdown solids concentration, short blowdown durations, or more precisely small blowdown volumes were used. Practically speaking, each blowdown volume should be no larger than the volume of the sludge thickening cone of the solids-contact unit. Given the desirable blowdown duration, the frequency of blowdown was dictated by the flux of solids through the unit. For high flow rates and sludge production, as with lime treatment or high carbon dosages, relatively frequent blowdowns were required.

Once a suitable blowdown duration and frequency was established by operating experience, adjustment of solids inventory was accomplished by increasing or decreasing the blowdown frequency. The proper solids inventory was obtained when the sludge and 5-minute sludge volume were at the levels desired. The blowdown frequency was then returned to approximately the previously established level.

During start-up of the chemical treatment unit, when solids are being accumulated, blowdown frequency was neven allowed to exceed more than two hours. This prevented accumulation of raw wastewater grit and probable plugging of the blowdown line.

Blowdown from the carbon contactors was commenced almost immediately after start-up. The reason for this operating procedure was that the desired carbon inventory was added to the unit prior to start-up.

Solids-contact treatment was defined as operation with the sludge level approximately 1/2 to 1 foot above the bottom of the reaction zone cone. This operation assured submergence of the draft tube inlet. When the sludge level was allowed to assume a depth greater than two to three feet above the bottom of the reaction zone cone, operation was defined as being solids-contact treatment with sludge-blanket clarification. With the type of solids-contact unit used in this study, the advantages of sludge-blanket clarification could be realized. This was because of the increasing area clarification zone which provided a more stable sludge level. Also of importance was the fact that all slurry to be clarified passed through the sludge blanket which existed above the bottom of the reaction zone skirt.

A summary of typical solids-contact unit operating conditions used in this study are presented in Table C-1.

TABLE C-1: SOLIDS-CONTACT UNITS OPERATING CONDITIONS

TREATMENT:	Chemical Contactor					Carbon Contactor	
	Lime pH 10.8	Lime pH 10.9	Lime pH 11.6	$FeCl_3$ 120 mg/ℓ	Alum 143 mg/ℓ	300 mg/ℓ	150 mg/ℓ
HYDRAULIC LOADING:							
gpm/sq ft	1.0	0.9	0.4	0.7	0.6	5-15	10-20
PUMPING TURBINE:							
rpm	45	37	23	29	12	30	30
Tip Speed, ft/sec	5.3	4.4	2.7	3.4	1.4	3.6	3.6
SLUDGE RAKE:							
rpm	0.56	0.80	0.56	--	0.26	0.30	0.30
Tip Speed, ft/sec	0.32	0.45	0.32	--	0.15	0.16	0.16
BLOWDOWN:							
Frequency, min	120	45	30	60	40	60	90
Duration, sec	10	20	15	25	30	21	10
Solids, g /ℓ	190	105	53	17	4	--	--
% Vol. of Total Flow	0.5	1.0	3.5	0.9	3.5	--	--
REACTION ZONE:							
Solids, g /ℓ	16	--	4	3	0.6	5-15	10-20
5-min Settled Vol., %	30	29	60	45	85	50	40
Sludge Pool Depth[a], ft	0	0	0	0	--	1-2	1/2-1
Approximate Sludge Flux, lb /MG	7000	7900	12000	1300	1300	>2500	<1250

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a - Depth above bottom of Reaction Zone Cone

APPENDIX D

GRANULAR MEDIA FILTER BACKWASHING OPERATION

The letters in the following procedure refer to Figure 19.

During the filtration mode of operation, the filter influent (A) and effluent (B) valves were open and the backwash outlet (C), air (AIR) and check valve (D) closed.

Backwashing the filter involved an automated sequence of valve munipulations and time delays. Backwash was manually or automatically initiated. Automatic backwashing was initiated when the filter bed headloss reached a preset (adjustable) value. The sequence of backwashing events were:

1. When the pre-set terminal headloss was reached (normally 7 ft of H_2O), the effluent sample pump, ΔP pressure tap solenoids and polymer feed pump were shut off, the influent valve (A) closed and backwash valve (C) opened.
2. An adjustable time delay (normally 25-30 minutes) allowed the liquid level in the filter to lower to within 2-4 inches of the filter bed surface.
3. The effluent valve (B) was then closed and the air valve opened.
4. Air scour for 0-10 minutes (adjustable, normally 1 minute) at 4-5 CFM/sq ft.
5. Air at 4-5 CFM/sq ft and water at 25 gpm/sq ft flowed simultaneously through the filter bed for about 1/2 minute until the liquid level in the filter neared the outlet at which time the air flow was shut off.
6. Hydraulic backwashing continued for 5-10 minutes (adjustable) after which the backwash pump shut off, backwash outlet valve (C) closed and influent valve (A) opened.
7. A 1/4 minute time delay was provided to insure that the rubber lined butterfly valve (D) had firmly seated, then effluent valve (B) opened allowing the water level in the influent stand pipe to assume an operating level of from 1/2 to 1 feet above the surface of the filter bed. At this time the effluent sample pump, ΔP sensor system and polymer feed pump were turned on.

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